ASSESSMENT OF PROCESS OPTIONS FOR TRITICALE FERMENTATION TO ETHANOL AND ANIMAL FEED IN THE WESTERN CAPE

by

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Declaration

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ABSTRACT

In South Africa there is a growing interest in the production of bioethanol for blending with petrol to reduce the environmental impact of fossil fuels. This project investigated the usage of triticale (small grains) for bioethanol production in the Western Cape (WC). Triticale is suitable for cultivation on marginal drylands in the WC. The project assumed that approximately 407 000 tonne/y triticale can be produced on these lands, allowing for construction and operation of a bioethanol-triticale plant with a production capacity of 160 Million ℓ ethanol/y.

Alternative process configurations for such a bioethanol facility were investigated in terms of energy balances and economic viability. This assessment included the conventional (warm) process, cold-hydrolysis process, warm pre-fractionation process and a combination of the cold and pre-fractionation processes. The following influences on the project's economic feasibly was investigated: A coal versus biomass fuel source, a combined-heat-and-power (CHP) plant option and external economic inputs.

The warm process is preferred over the cold process, since it has a higher Internal Rate of Return (IRR) (3.02% versus 2.40%). The warm process is also preferred above the warm pre-fractionated process as again the warm process gives a higher IRR. The pre-fractionated process produced less Dried Distillers Grains and Solubles (DDGS) containing a higher protein content, which can be sold at a higher price. To make the pre-fractionation process more profitable, the selling price of the pre-fractionated DDGS should be between 2.5-4 times higher than the DDGS without pre-fractionation.

The use of biomass as fuel source for energy rather than coal is recommended, since it is less expensive in the WC. Biomass reduces the carbon emissions of the process by 25%.

The project recommended the use of a CHP plant for onsite steam and electricity production with sales of surplus electricity to nearby users. The Capital Expenditure (CAPEX) of the plant increases with 30% when using CHP, but this increase is mitigated by the selling of excess electricity.

The Basic Fuel Price (BFP) and triticale price predominantly influence the plant's profitability. Therefore, the calculations of government subsidy for plant should be dynamic, and the subsidy should be revised monthly in accordance with the BFP and triticale price variations.

The current 15% Return on Assets (ROA) subsidy calculations was insufficient to achieve an expected project IRR of 9.7%, under base case conditions. Therefore, an alternative subsidy mechanism needs to be investigated, or a significantly lower

triticale grain price should be sought.

Using sorghum as the reference grain for a triticale ethanol production plant has a large effect on IRR. A triticale grain price significantly below SAFEX B4 wheat and SAFEX sorghum is essential for a bioethanol plant to be economically viable. Therefore, a detailed market analysis needs to be done for triticale and DDGS prices (prices should be secured by a contract).

It is recommended that all processes should be tested on lab and pilot plant scale. In conclusion, this project recommends the warm process with a CHP plant using biomass as fuel for energy source for ethanol production from triticale.

OPSOMMING

Daar is in Suid-Afrika ‘n toenemende aanvraag na die produksie van bio-etanol in die vermenging van petrol om fossielbrandstowwe se omgewingsimpak te verminder. Korog (kleingrane) is in hierdie projek in die Wes-Kaap vir bio-etanolproduksie ondersoek. Marginale droëlande is geskik vir Korog verbouing in die Wes-Kaap. Die projek het aanvaar dat ongeveer 407 000 ton/jaar korog op hierdie lande geproduseer kan word, wat aanleiding gegee het dat ‘n produksiekapasiteit van 160 Miljoen ℓ etanol/jaar korog bio-etanol-aanleg gebou en bedryf kan word.

Vir die bio-etanol-aanleg is alternatiewe proses konfigurasies, in terme van energiebalanse en ekonomiese lewensvatbaarheid, ondersoek. Hierdie ondersoek het die gewone (warm)proses, koue-hidroliseproses, warm pre-fraksioneringsproses en ‘n kombinasie van die koue en pre-fraksioneringsprosesse ingesluit. Die volgende invloede op die projek se ekonomiese volhoubaarheid is ondersoek: ‘n Steenkool teenoor biomassa brandstofbron, ‘n gekombineerde warmte en krag (GWK) aanleg en eksterne ekonomiese insette.

Die warmproses word bo die koueproses verkies, aangesien dit ‘n hoër interne opbrengskoers (IOK) het (3.02% teenoor 2.40%). Die warmproses word ook bo die warm pre-fraksioneringsproses verkies, aangesien die warmproses weereens ‘n hoër IOK het. Die pre-fraksioneringsproses produseer minder GDGO (Gedroogte distilleerde graan en oplosbarestowwe) maar met ‘n hoër protein-inhoud wat dus teen ‘n hoër prys verkoop kan word. Om die pre-fraksioneringsproses winsgewend te maak moet die pre-fraksionerings GDGOverkoopprys tussen 2.5 en 4 keer hoër wees as dié van GDGO sonder pre-fraksionering.

Biomassa eerder as steenkool word aangebeveel as energie brandstofbron, aangesien dit goedkoper in die Wes-Kaap is. Biomassa verminder die projek se koolstofvrystelling met 25%.

Die gebruik van ‘n GWKaanleg word aanbeveel, om stoom en elektrisiteit op die perseel te vervaardig, waarna die oortollige elektrisiteit aan nabygeleë verbruikers verkoop kan word. Die kapitaalkoste verhoog met 30% as GWK gebruik word, maar die effek op IOK word teengewerk deur die verkoop van oortollige elektrisiteit.

Die aanleg se winsgewendheid word hoofsaaklik deur die basiese brandstofprys (BBP) en korogprys beïnvloed. Dus moet die subsidie berekeninge dinamies wees en derhalwe maandliks hersien word na aanleiding van die BBP en korogprys variasies.

Die huidige 15% ondernemingsrentabiliteit subsidie was onvoldoende, aangesien die projek se verwagte IOK van 9.7% nie onder die huidige omstandighede bereik kon word nie. ‘n Alternatiewe subsidiemeganisme moet dus ondersoek word of

alternatiewelik moet daar gepoog word om ‘n aansienlike laer korogprys te bekom.

Sorghum as ‘n verwysingsgraan vir ‘n korog etanolproduksie-aanleg het ‘n groot invloed op die aanleg se IOK. ‘n Korogprys wat aansienlik laer as die SAFEX B4 graan en SAFEX sorghumprys is, is noodsaaklik vir die ekonomiese lewensvatbaar van ‘n bio-etanol-aanleg. ‘n Omvattende mark analise moet dus op die korog en GDGOpryse (pryse behoort deur ‘n kontrak verseker te word) gedoen word.

Daar word aanbeveel dat alle prosessse op laboratorium en proefaanlegskaal getoets moet word. Ten slotte, die warmproses met ‘n GWKaanleg wat biomassa as energie brandstofbron gebruik vir etanolproduksie word deur die projek aanbeveel.

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ABBREVIATIONS

ABBREVIATION	WORD
BBP	Basiese Brandstofprys
BFP	Basic Fuel Price
CAPEX	Capital Expenditures
CHP	Combined Heat and Power
CBP	Consolidated bioprocessing
CEPCI	Chemical Engineering's Plant Cost Index
CO_2	Carbon Dioxide
CSTR	Continuous Stirred Reactor
DDGS	Dried Distillers Grains and Solubles
FAN	Free Amino Nitrogen
FCI_L	Fixed Capital Investment
EBITDA	Earnings before Interest, Tax, Depreciation and Amortisation
EU	European Union
GDGO	Gedroogte distilleerde graan en oplosbarestowwe
GHG	Greenhouse Gases
GWK	Gekombineerde Warmte en Krag
IAP	Invasive Alien Plants
IOK	Interne Opbrengskoers
IRR	Internal Rate of Return
JSE	Johannesburg-based exchange
LCA	Life Cycle Assessment

LHV	Lower Heating Value
MACRS	Modified Accelerated Cost Recovery System
MESP	Minimum Ethanol Selling Price
MTBE	Methyl-tert-butyl ether
NPV	Net Present Value
OPEX	Operational Expenditures
PH	Process Heat
ppm	Parts per million
ppb	Parts per billion
ROA	Return on Assets
REIPPPP	Renewable Energy Independent Power Producer Procurement Programme
SA	South Africa
SAFEX	South African Futures Exchange
SHF	Separate Hydrolysis and Fermentation
SSF	Simultaneous Saccharification and Fermentation
USA	United States of America
Vs.	Versus
WC	Western Cape
WDG	Wet Distillers Grains
wt	mass/weight
WWII	World War II

NOMENCLATURE

SYMBOLS	DESCRIPTION	UNITS
C_1	Capital cost plant 1	R
C_2	Capital cost plant 2	R
$C_{2\ corrected}$	Capital cost plant 2 corrected for inflation	R
Cv	Calorific value	kWh/kg
h_w	Enthalpy of water	kJ/kg
I_1	Historically CEPCI value	1
I_2	Present CEPCI value	1
IRR	Internal Rate of Return	%
i	Interest rate	%
LHV	Lower Heating Value	kJ/kg
m	Mass	kg
N	Lifespan of plant	year
NPV	Net Present Value	R
n	Year	year
η_{eff}	Boiler efficiency	1
Q_1	Ethanol capacity 2	ℓ/year
Q_2	Ethanol capacity 2	ℓ/year
ROA_{max}	Return on Assets for constant asset value subsidy method	%
$ROA_{straight}$	Return on Assets for depreciated asset value subsidy method	%
Sub_{max}	Subsidy for constant asset value subsidy method	R/ℓ
$Sub_{straight,\ nominal}$	Nominal subsidy for constant asset value subsidy method	R/ℓ
$Sub_{straight,\ \mathrm{real}}$	Real subsidy for constant asset value subsidy method	R/ℓ
X	Scale up factor	1

THESIS

1 INTRODUCTION

The key objective of this project is to simulate the production of bioethanol from triticale. Different process configurations or scenarios will be evaluated to determine which process is the most economically feasible for the Western Cape (WC). Discussed below are the background of the project, key questions and the hypothesis, deliverables and the scope of the project.

1.1 Background

Environment pollution caused by fossil fuels, specifically in road transportation, is a growing concern (Balat and Balat, 2009). The carbon dioxide (CO_2) produced from fossil fuel combustion can lead to climate change. Renewable fuels such as bioethanol and biodiesel may contribute to reducing CO_2 emissions, which in turn can reduce the impact on the environment (Balat and Balat, 2009).

Carbon dioxide from the atmosphere is transformed into plant biomass via photosynthesis. This biomass can be used in the production of bioethanol. As a replacement for fossil fuels, the production and use of bioethanol is likely to lead to a reduction in net CO_2 emissions (Balat and Balat, 2009).

In comparison to biofuels, the use of fossil fuels does not provide such a “closed carbon cycle,” but rather results in a net increase in Greenhouse Gas (GHG; e.g. CO_2) concentrations in the atmosphere (Balat and Balat, 2009). Fossil fuels are furthermore likely to become scarcer and more expensive in the long term, whereas biofuels are inherently renewable and sustainable (Balat and Balat, 2009). Bioethanol is cleaner-burning than fossil fuels as it has a low sulphur and heavy metal content (Balat et al., 2008). It is an oxygenated fuel and therefore complete combustion takes place resulting in less carbon monoxide being emitted (Balat et al., 2008).

Currently South Africa (SA) produces very little bioethanol for use as a biofuel .This is most likely due to concerns about their economic feasibility, increasing food prices, and the food versus fuel debate (Pradhan and Mbohwa, 2014). Current bioethanol production, using conventional feedstock produced on arable land, is focused exclusively on the potable and beverage grade markets (Pradhan and Mbohwa, 2014). There is, however, the potential to expand bioethanol production in the WC through the cultivation of triticale as feedstock, using marginal lands (Melamu, 2015). Marginal drylands are lands that are no longer used for wheat production (Melamu, 2015). Bioethanol production can be increased when sufficient subsidy support is available from the national government to ensure the economic viability of biofuels

production.

Triticale is a hybrid between wheat and rye and is presently used as an animal feed and ground cover in WC agriculture. It is a viable biofuel crop, especially because it can be cultivated on marginal lands, contains high starch levels and is suitable as fermentation feedstock. The drylands area under small grains cultivation has decreased significantly in recent decades (Melamu, 2015). This is due to changing market conditions and a preference for high-yielding soils. Large areas of drylands are no longer under wheat cultivation as the wheat yields on these are not profitable (Melamu, 2015). Triticale, however, have the advantages that the grain is drought tolerant, pest and disease resistant, needs less input cost and produces a higher yield (Kučerová, 2007). It is therefore a more profitable crop to grow (Kučerová, 2007). Another advantage is that triticale production on marginal lands will limit the potential impact of biofuels on food production.

For the purpose of this project it was assumed that 932 000 tonne triticale grains can be produced per year by means of crop rotational cycles with wheat on non-marginal lands in conjunction with the use of marginal lands (Melamu, 2015). As a result the production capacity for bioethanol-triticale of more than 160 million litres of ethanol per year can be produced (Melamu, 2015). It is estimated that the amount of hectares of marginal lands in the WC is 70 000 ha and if crop rotation cycle are included 381 000 ha (Melamu, 2015). Using triticale in crop rotation with wheat can improve the yields of wheat production as triticale replenishes nutrients in the soil. The average yield of triticale per hectare is 2.5 tonnes (Melamu, 2015).

Triticale produced in the WC for ethanol production and other local uses will avoid the “transport differential” penalty. This penalty of approximately R600/tonne is paid by local farmers for the “export” of grains to inland markets (national grain prices are determined in Gauteng) (Coetzee, 2015). The agricultural potential of triticale as a biofuel crop therefore warrants further investigation to determine the economic and environmental potential of triticale-ethanol production plants in the WC.

1.2 Key questions and hypothesis

The main hypothesis is:

- Triticale can be used as feedstock for bioethanol production in a way that it is economically feasible in the Western Cape.

The key questions discussed are as follows:

- Is it more cost effective to use the warm or the cold process for liquefaction of starch?
- Does pre-fractionation make the dry grinding process more profitable?
- Does the profitability of the process increase if a Combined Heat and Power (CHP) plant is used for onsite steam and electricity production, rather than only producing steam onsite through a low-pressure boiler and purchase-in of process electrical demands from the national grid?
- How does changing the energy fuel source to generate steam, from coal to biomass, decrease the process profitability?
- Which of the input parameters to the economic model have the greatest influence on the profitability of the process, based on sensitivity in the model outputs and historical variations in these inputs?
- Is the method of subsidy calculation proposed by national government sufficient to ensure the profitability of the process?

1.3 Deliverables

The following deliverable will be achieved from this project:

- Project Proposal;
- Master's Thesis (this report);
- Master's Oral;
- Four Aspen Models; and
- Sixteen Economic Models.

1.4 Scope

In this project, triticale is used as feedstock to produce bioethanol. Different process variations and scenario are compared to determine which process is the most economically feasible. The process models are the (1) warm process, (2) cold process, (3) warm pre-fractionation process as well as (4) a combination of the cold and pre-fractionation processes.

The option for onsite electricity and steam production in a CHP plant, in comparison to the combination of a low-pressure onsite boiler for steam production and buying in electricity from Eskom, was also considered. The effect of coal as opposed to biomass as the fuel source for steam/electricity generation on the profitability of the project is furthermore investigated for different process models. An economic feasibility analysis is done for all of these processes. Additionally, different ways to calculate the subsidy based on the Return on Assets (ROA) are investigated.

A sensitivity analysis was done to determine how different fluctuations in triticale price, Basic Fuel Price (BFP), Dried Distillers Grains and Solubles (DDGS) price, CO_2 price, biomass price, coal price, Capital Expenditures (CAPEX) and capacity influence the profitability of the process. Additionally, a historical price analyses was done for BFP and triticale prices from January 2009-April 2015 to determine if the process was profitable. A CO_2 balance was done for the model that achieved the highest Internal Rate of Return (IRR). The balance was done to compare the effect of using biomass rather than coal on the carbon balance.

A literature study in regard to bioethanol production was done in the following section. This literature will be used as the working foundation for this project.

2 LITERATURE REVIEW

The literature that is applicable to this project is discussed below. The literature in this section starts with the historical overview of ethanol and its role throughout history.

2.1 Historical Overview

For more than a millennium, ethanol has been used as an alcoholic drink, but only from 1850 onwards has ethanol been used as a lightning fuel (Abebe, 2008). In 1824-1826, Samuel Morey developed the first combustion engine. It used a mixture of turpentine and ethanol as fuel (Ethanolhistory, 2011). During the American Civil War (1861-1865) in the United States of America (USA) liquor tax was levied on ethanol to raise funds for the war. Consequently, ethanol became non-competitive with other lighting fuels like kerosene and ethanol production levels decreased drastically. When the tax was removed in 1906, ethanol production levels again started to rise (Ethanolhistory, 2011).

In 1897, ethanol was used as fuel by Nikolas Otto for the internal combustion engine (Bowonder, 1983). Henry Ford build his first automobile in 1896, the Quadricycle, which used ethanol exclusively as fuel (Ethanolhistory, 2011). Later on Henry Ford also designed his model T Ford to run on a mixture of bioethanol and gasoline in the early 1900s (Abebe, 2008). In the 1920s, Brazil received its first motor vehicles for which they used ethanol as its fuel source(Ethanolhistory, 2011). When prohibition started in 1919 in the USA ethanol was banned. It could only be sold when it was mixed with gasoline. At the end of prohibition in 1933, ethanol was once again used as fuel and consequently its consumption level increased. During World War II (WWII) ethanol consumption significantly increased as oil became scarce, bringing about a demand for an alternative fuel. Ethanol consumption, however, again diminished after WWII (Abebe, 2008).

In 1970 the major oil producing countries cut the amount of gasoline they supplied, instigating a renewed interested in bioethanol production in the USA (Bothast and Schlicher, 2005). Bioethanol production was encouraged by offering tax incentives to lessen the USA‘s dependence on foreign oil and stimulate agricultural growth (Balat and Balat, 2009). In 1974, the USA enacted the Solar Energy Research, Development, and Demonstration Act of 1974. Brazil followed a year later with its own Programa Nacional do Álcool to encourage and research the production and use of alternative organic fuels (Ethanolhistory, 2011). In 1979 Amoco Oil Company started mixing ethanol with gasoline and their example was soon followed by other oil companies (Ethanolhistory, 2011).

When ethanol was added as oxygenate in 1988 to lessen carbon monoxide emissions, ethanol production increased. The Clean Air Act of 1990 in the USA

aimed to remove toxins like benzene, toluene and zylene (oxygenates) from gasoline to create cleaner emissions. When the alternative oxygenate MTBE (Methyl-tert-butyl ether) was banned in 2000, since it was contaminating ground water, the demand for bioethanol increased again. The Energy Policy Act of 1992 in the US aimed to promote vehicles that run on alternative fuels (e.g. E85) through tax incentives (Ethanolhistory, 2011). Brazil was the top producer of bioethanol until 2005, after which the USA started producing more bioethanol than Brazil (Balat et al., 2008). See Figure 1 for historical timeline.

From the above discussion, the production of ethanol can be seen as a growing industry. It is a more environmentally friendly alternative to gasoline, which is currently used as a major fuel source (Cai et al., 2013; Kaliyan et al., 2011; M. Wang et al., 2007). An in-depth discussion of bioethanol continues in the next section.

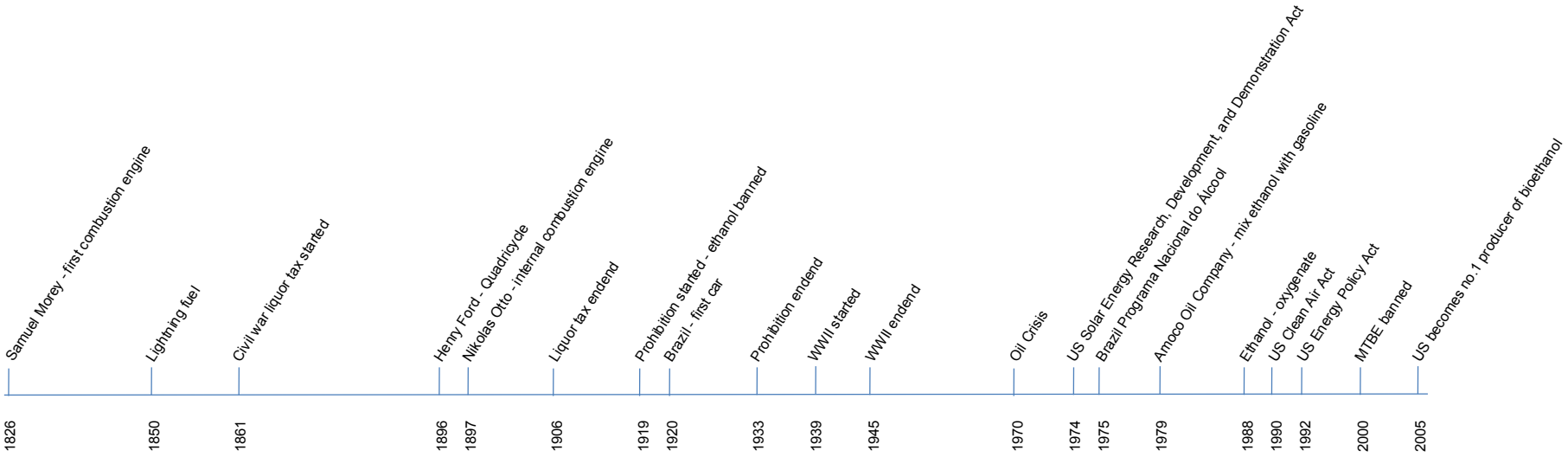

FIGURE 1: HISTORICAL TIMELINE

2.2 Bioethanol

Bioethanol can be used in isolation or in a mixture (primarily with gasoline) as a fuel for use in motor vehicles. This depends if the engine of the motor vehicle is modified or not. Ethanol has a higher oxygen (35%) content than gasoline, resulting in a corresponding decrease in energy content. Less oxygenate additives such as MTBE need to be added to achieve the desired oxygen content of fuel for clean combustion (Bothast and Schlicher, 2005). The increased oxygen content in the fuel facilitates the process of complete combustion. Therefore, fewer detrimental (toxic) products such as carbon monoxide, NO_x and aromatic compounds are produced. More non-toxic compounds, such as CO_2, are produced in their place.

Carbon monoxide is more detrimental to the environment than CO_2, due to it being toxic. Ethanol is also non-toxic and therefore it does not contaminate water sources as in the case of gasoline (Balat et al., 2008). The CO_2 produced from burning bioethanol is carbon neutral. The reason for this is that the amount of CO_2 released from burning ethanol is the same as the amount of CO_2 consumed by the growth of the plant/feedstock to produce bioethanol (Balat and Balat, 2009).

Ethanol (108) also has a higher octane number than gasoline (87-93) (Federal Trade Commission, 2012; MacLean and Lave, 2003). A higher octane number helps prevent early ignition, which can cause cylinder knocking. Ethanol has a higher flame speed and broader flammability limits than gasoline and vaporises at a higher temperature. The above properties of ethanol lead to a higher compression ratio, leaner burn engine and shorter burn time. In comparison to gasoline, ethanol is the more beneficial choice with regard to engine performance and the environment.

Bioethanol; however, also has disadvantages. Its energy density is lower than gasoline's and it only produces 66% of the energy gasoline does. Bioethanol is also corrosive, has a lower flame luminosity and a lower vapour pressure, which may cause difficulties when starting a motor vehicle at low temperatures (MacLean and Lave, 2003). Bioethanol is mostly blended to gasoline in a 10% ratio. This is called E10 in the US. E10 consist of the maximum amount of ethanol motor vehicles can run on without modification. Motor vehicles that are modified to use ethanol are called Flexfuel motor vehicles and can use E85, which is 85% ethanol and 15% gasoline (Demirbas, 2008).

The advantages for using ethanol, as opposed to gasoline, outweigh the disadvantages.

Bioethanol is even more advantageous than ethanol. This is the case, since it refers to ethanol being produced by a biological process. Ethanol can be produced either synthetically or biologically depending on the process used to create it. Biologically produced ethanol is more environmentally friendly than synthetically produced

ethanol. The classification of whether ethanol is produced biologically or synthetically depends on the feedstock used in its production.

2.3 Feedstocks

Bioethanol production can be classified by the feedstock that is used for production, namely first generation (sucrose and starch), second generation (lignocellulosic) and third generation (algae, municipality waste) feedstocks. The three most common types of feedstocks used to produce ethanol are sucrose-containing feedstocks, starch-containing feedstocks and lignocellulosic biomass. These three types of feedstocks are discussed in more detail below. .

2.3.1 Sucrose-containing feedstocks

The primary sucrose-containing feedstocks consist of the following crops, namely sugar cane, sugar beet and sweet sorghum. Sugar cane offers the advantages of a high yield in sugar per hectare and a low conversion cost for processing sugar (Vohra et al., 2013). Therefore, less operation cost is needed to operate a sugar cane plant in comparison with other types of plants, as less feedstock is needed and the conversion from sucrose to glucose is easy (Balat et al., 2008; Cardona and Sánchez, 2007). In contradiction to this Pradhan (2014) states that in a sugar cane ethanol production plant in SA is more expensive than a sorghum ethanol plant. Therefore, the use of sugar cane in comparison with other grain feedstocks in bioethanol production can be more expensive in SA than internationally.

A disadvantage of sucrose-containing feedstocks is that they are seasonal and therefore plants cannot operate all year round (Karhammar et al., 2006). Most of the world's bioethanol that is produced from sucrose-containing feedstocks is produced from sugar cane (Dhavala et al., 2006).

Sugar cane is grown in subtropical and tropical climates and therefore it is not suited for the more temperate climates that is typical of the WC. Most of SA has very low, low or medium rainfall patterns that is not suitable for sugar cane production (Department of Minerals and Energy, 2007). Consequently, no significant expansion (only ± 45 000 ha) of sugar cane plantations is feasible in SA. Therefore, sugarcane is not a viable option for ethanol production on large scale in SA (SASA, 2015). To use sugar cane as a feedstock the biofuel industry needs to compete with the food industry as sugar cane is used in human consumption. The biofuel industry is therefore in competition with the human food consumption industry for feedstock (Meyer et al., 2005; Pradhan and Mbohwa, 2014). In other words, the high price of feedstock and the food versus fuel debate has a negative impact on the suitability of using sugar cane in the production of ethanol.

The food versus fuel debate arose out of concerns about food security (Lemmer and

Schoeman, 2011). If food is used for fuel production it can increase the cost of food since these different markets have to compete for the same commodity (Richard et al., 2012). There is also a debate on the immorality of using food for fuel production while there are people starving in the world (Richard et al., 2012). It is also reasoned that fuel is needed to ensure the proficient production of food and therefore a decrease in fuel price would result in lower food prices (Richard et al., 2012). Furthermore, food prices are dependent on climate conditions and therefore if climate change is not partly alleviated by using biofuels, eventually the sustainability of food production will affected (Richard et al., 2012).

Other types of sucrose-containing feedstocks that can be used are sugar beets and sweet sorghum. Sugar beets are mostly grown in European countries. Sugar beets require less water and fertilisation than sugar cane and are therefore a good alternative for bioethanol production. Currently it is proposed that a sugar beet ethanol plant be built near Cradock in the Eastern Cape (Nasterlack et al., 2014). Sweet sorghum is also a very promising sucrose containing feedstock, but it is not widely used as the crop is still under development (Dhavala et al., 2006).

2.3.2 Lignocellulosic biomass feedstock

Lignocellulosic biomass feedstocks consist of agricultural residues, hard/soft wood, cellulose waste and herbaceous biomass (Balat et al., 2008; Vohra et al., 2013). Lignocellulosic biomass contains lignin, cellulose and hemicellulose. Cellulose fibres consist of hexose sugars (glucose, etc.) and are easily fermentable. Hemicellulose consists mainly of pentose sugars (xylose, etc.), which are more difficult to ferment. Most preferred industrial fermentative yeasts cannot ferment pentose or if they can, they achieve low ethanol yields. Research is currently being done to genetically modify organisms in order to be able to ferment both hexose and pentose with equal ease and hence increase the ethanol yield of the process (Balat et al., 2008).

Lignocellulosic biomass is usually very cheap as it has a low or no market value. The only costs usually involved are transport costs. This is an advantage, because feedstock can comprise up to 40% of the process cost (Govindaswamy and Vane, 2007). Agricultural residues in the WC are left on the fields as part of conservation agriculture. This is done to conserve the moisture and nutrient content in soil (Melamu, 2015). Due to this already established use of residues, it is not a feasible option as a feedstock for bioethanol production.

Invasive Alien Plants (IAP) is available as a feedstock for bioethanol production as it has currently very low market value (Nowell, 2011). The cost involved in producing bioethanol from biomass is relatively high due to high cost of hydrolysis and low ethanol yields (Balat et al., 2008). Moreover, additional pre-treatment is needed to make the biomass susceptible for hydrolysis and fermentation (Sánchez and Cardona, 2008). This increases the capital and operational cost of the process

(Cardona and Sánchez, 2007). The lignocellulosic biomass process has the potential to be profitable in the future, but at present more research is needed to improve the process.

2.3.3 Starch-containing feedstocks

Starch-containing feedstocks like corn, wheat, barley, sorghum, triticale, rye and cassava are suitable for industrial ethanol production. Starch consists of 1000-6000 monomers that are linked together to form a homopolymer (Vohra et al., 2013). This homopolymer contains only D-glucose monomers (Balat et al., 2008). Starch consists of two main structures, namely amylopectin and amylose. Amylopectin consists of 4% α-1,6 linkages of glucose molecules (branched) and 96% α-1,4 linkages of glucose molecules (unbranched) and has a molecular weight of 50-200 million g/mol. Amylose in comparison only has 0.1% α-1,6 linkages and has a molecular weight of 200 000-700 000 g/mol. Typically, 70-80% of the starch consist of amylopectin (Smith et al., 1997). Usually high yields of ethanol are obtained when starch is fermented due to it consisting of glucose. Table 1 below summarize the starch content an liquefaction temperature of different starch-containing feedstocks.

TABLE 1: PROPERTIES OF DIFFERENT STARCH-CONTAINING FEEDSTOCKS

Starch-containing feedstocks	**Corn**	**Wheat**	**Barley**	**Sorghum Grain**	**Rye**	**Trriticale**
Starch Content	65% wt[1]	65% wt[2]	50-55% wt[3]	64-74% wt[4]	60-63% wt[5]	66% wt[6]
Liquefaction Temperature	90°C[7]	65°C[8]	-	86°C[9]	80°C[10]	60°C[11]

[1] From (Gago et al., 2013)
[2] From (Gago et al., 2013)
[3] From (Hicks et al., 2005)
[4] From (Wang et al., 2008)
[5] From (Wang et al., 1997)
[6] From (Pejin et al., 2009)
[7] From (Gago et al., 2013)
[8] From (Gago et al., 2013)
[9] From (Wang et al., 1999)
[10] From (Wang et al., 1999)
[11] From (Pejin et al., 2009)

The different types of starch used in fermentation are elaborated on below in the below sections.

2.3.3.1 *Corn*

Corn is the primary starch based feedstock used to produce bioethanol. Ninety percent of the bioethanol produced in the USA comes from corn (Balat et al., 2008). The USA is currently the country that produces the most bioethanol per year (Balat et al., 2008). Corn contains about 65% wt starch (Gago et al., 2013). The conventional dry grinding liquefaction temperature for corn is 90°C (Gago et al., 2013). It is also used for food production and hence the fuel versus food debate resurfaces (Patni et al., 2013).

2.3.3.2 *Wheat*

Wheat grains have a starch content of ± 65%, similar to that of corn. The most important difference between wheat and corn is wheat's lower gelatinisation temperature. Therefore, the conventional dry grinding liquefaction temperature for wheat is 65°C, while for corn it is 90°C (Gago et al., 2013). Wheat is used as a food source and therefore using wheat to produces bioethanol invokes the fuel versus food debate (Patni et al., 2013).

2.3.3.3 *Barley*

To produce bioethanol from barley, the conventional dry grinding process needs to be modified. This affects the profitability of the process and makes barley less economical feasible than corn or wheat (Hicks et al., 2005). The advantage of barley above corn is that it can be grown in less desirable circumstances (Hicks et al., 2005). Barley has an abrasive hull that damages the grain-handling and grinding equipment and therefore increases the capital cost of the process. Furthermore, barley also has a low starch content of between 50% and 55%. Hence, to render the process cost efficient the starch content of the barley needs to be increased (Hicks et al., 2005). In addition, barley mash has a high viscosity that makes mixing, pumping and fermenting difficult and therefore the high viscosity causes operating costs to be high (Hicks et al., 2005). Further research is being done to create hull-less, high-in-starch content varieties of barley for bioethanol production. This should reduce both the operational and capital costs of the process. One of these varieties is Doyce, which is produced by Virginia Tech (Hicks et al., 2005). Commercial β-glucanase enzymes can be added to the barley mash to decrease the viscosity and therefore the Operational Expenditures (OPEX) of the process (Hicks et al., 2005).

2.3.3.4 *Sorghum grain*

Sorghum grain has a starch content of 64-74% (Wang et al., 2008). Liquefaction for

sorghum is done at 86°C, which is higher than for wheat and triticale (Zhao et al., 2009). Sorghum is also drought resistant and can produce high amounts of grain under non-ideal conditions. In general, waxy and heterowaxy sorghum varieties produce higher ethanol yields than non-waxy sorghum varieties. This is because waxy and heterowaxy sorghum varieties contains mostly amylopectin (Shewale and Pandit, 2009; Wang et al., 2008). Sorghum varieties with tannin usually have lower starch hydrolysis, as tannin acts as an inhibitor (Wang et al., 2008). Sorghum that contains tannin has a higher mash viscosity and therefore the efficiency of fermentation decreases, as there is a negative correlation between mash's viscosity and fermentation efficiency (Wang et al., 2008).

2.3.3.5 *Rye*

Rye (60-63%wt) has a lower (2-5% lower) starch content than wheat but is less expensive to procure (Wang et al., 1997). It also has a higher content of hemicellulose than other grains and therefore it forms a viscous mixture in mashing (Wang et al., 1997). Therefore enzymes (Roxazyme G) needs to be added to mitigate this effect (Wang et al., 1997). Adding urea into the mixture shortens the fermentation time with 40% for rye, from 120 hours to 72 hours (Wang et al., 1998). This is shortening is due to the high level of Free Amino Nitrogen (FAN) in the grain (Wang et al., 1998). Liquefaction is done at 80°C for rye (Wang et al., 1999).

2.3.3.6 *Triticale*

Triticale is a hybrid between wheat and rye (Tsupko, 2009). It is mostly used as an animal feed and ground cover in agriculture and consequently there is no competition with human food (Tsupko, 2009). The starch content for the Odyssey triticale variety is 66.35% starch and 12.65% protein on dry matter (Pejin et al., 2009). Triticale has the following advantages over wheat and rye (Kučerová, 2007):

- It produces a higher grain yield even under unfavourable conditions;
- It is resistant to soil-climatic conditions (NST 21/06);
- It has a tolerance to dryness;
- It has a higher tolerance to acid soils;
- It requires a lower amount of nutrient substances;
- It is more pest- and disease-resitant.

As a result triticale needs less fertilisation and chemicals to produce a good grain yield, making it a more profitable crop to grow (Kučerová, 2007).

Some triticale varieties contain native enzymes, which means that no extra enzymes need to be added for complete hydrolysis (Pejin et al., 2009).

In conclusion, it is more profitable to use triticale than wheat for ethanol production, as the feedstock price is lower due to lower agricultural input cost. This increased profitability could maybe also be ascribed to the fact that it may not require enzymes, which also lowers the cost of ethanol production.

It is evident from the different feedstocks used for bioethanol production considered above, that the advantages of triticale outweigh those of the other feedstocks. One of the most important advantages is that triticale is not a competing factor in the human consumption industry, which considerably lowers the cost of production. The hydrolysis and fermenting method of the feedstock; however, needs to be taken into consideration as it will affect the enzyme efficient and the ethanol yield. This in turn could affect the profitability of the plant.

2.4 Hydrolysis and Fermenting Technologies

Different types of processes and combinations of processes are used during hydrolysis and fermentation. In this section there are four important processes/concepts they are liquefaction, saccharification, hydrolysis and fermentation. Liquefaction is the process where starch is liquefied from solid in into soluble liquid (reduce dextrin chain length) by the enzyme α-amylase (Novozymes, 2016). Saccharification is when the liquefied starch is converted into glucose by glucoamylase (Biology-Online, 2015). Hydrolysis refers to the enzymes action of breaking bonds in combination with water addition (Biology-Online, 2015). Fermentation refers to the process where the glucose (sugar) is converted into ethanol and CO_2 by saccharomyces cerevisiae (yeast) (Biology-Online, 2015).

See Figure 2 for the different processes:

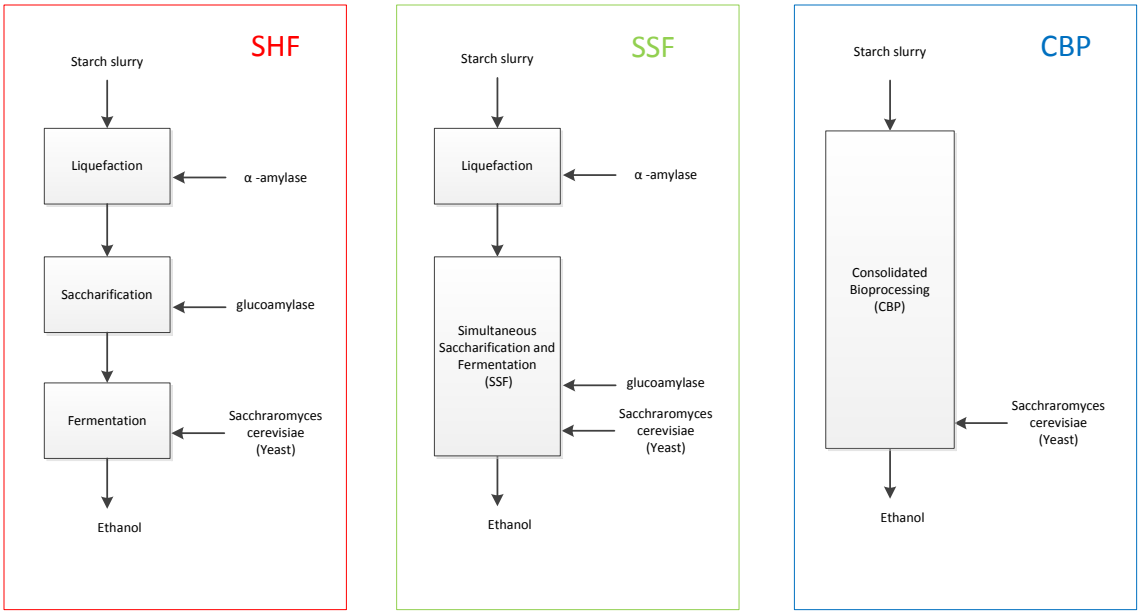

FIGURE 2: PROCESS DIAGRAMS OF SEPARATE HYDROLYSIS AND FERMENTATION (SHF), SIMULTANEOUS SACCHARIFICATION AND FERMENTATION (SSF) AND CONSOLIDATED BIOPROCESSING (CBP)

REDRAWN FROM: (Mojovic et al., 2009; Van Rensburg et al., 2013)

2.4.1 Separate hydrolysis and fermentation (SHF)

SHF happens when hydrolysis and fermentation take place in different reactors; see Figure 2. Here the hydrolysis and fermentation may ensue under the optimal conditions for each (Balat et al., 2008). Enzymes operate optimally at 70°C for triticale, while the optimal temperature for yeast is 35°C (Pejin et al., 2009). It should be noted that glucose inhibition starts to take place if the glucose concentration is too high and therefore even if not all the glucose is converted, no more glucose will be produced (Balat et al., 2008; Mojovic et al., 2009). In addition, the risk of contamination increases as sugar is not immediately used and other organisms can start to grow on the sugars (Mojovic et al., 2009).

2.4.2 Simultaneous saccharification and fermentation (SSF)

SSF occurs in one reactor where both hydrolysis and fermentation take place (Figure 2). Enzymes and yeast are usually added to the same reactor and as the enzymes hydrolyse the starch the yeast converts glucose to ethanol. The operating conditions for this type of reactor are 35°C with a pH of 4.8 (Balcerek and Pielech-Przybylska, 2013). The advantage of this type of setup is that there is no glucose inhibition, as glucose is used as it is produced (no glucose build-up). The risk of contamination increases as the reactor is operated at lower temperatures which favour micro-organisms (Mojovic et al., 2009). This contamination is mitigated by allowing the

yeast to use glucose as it is produced and hence only small amounts of sugars are available to be used by other organisms. Furthermore, the ethanol present in the fermentation broth also decreases the risk of contamination as it kills most of the bacteria (UCSB Science Line, 2015). The disadvantage of this process is that the enzymes do not operate at their optimal conditions and therefore the rate of production of glucose decreases (Cardona and Sánchez, 2007). Surprisingly, even though the enzymes do not operate at their optimal conditions, this type of process has a higher ethanol yield than SHF due to the glucose inhibition. It is therefore preferred industrially to SHF (Mojovic et al., 2009).

The capital cost of SSF is lower than that of SHF, as one reactor instead of two reactors can be used to achieve the same results (Balat et al., 2008). Currently research is being conducted to make the fermenting organisms tolerant of higher temperatures, which might improve the enzymatic yield of this process (Cardona and Sánchez, 2007).

2.4.3 Consolidated bioprocessing (CBP)

CBP is similar to SSF in that both hydrolysis and fermentation occur simultaneously in a single reactor vessel, although with CBP no external enzymes are added to the process, as the required enzymes are produced by the yeast (Vohra et al., 2013), see Figure 2. Yeast can be genetically modified to produce enzymes for hydrolysis, which will consequently reduce the cost associated with buying enzymes. CBP has lower ethanol yields as yeast is inoculated resulting in low enzyme production (Van Rensburg et al., 2013). Therefore, small amounts of starch are converted to glucose, which means that small amounts of glucose are available for yeast usage. Hence, as the yeast grows, more and more enzymes will be produced and thus more starch is hydrolysed to glucose (Van Rensburg et al., 2013).

The problem with this type of process is that it does not produce enough enzymes in the beginning of the process and hence the fermentation takes longer (Van Rensburg et al., 2013). In some instances it also does not produce enough enzymes to hydrolyse the starch and therefore the ethanol yield is lower (Nkomba, 2015). A way to mitigate this effect is to add additional enzymes at the beginning of fermentation to convert starch to glucose. The size of the yeast inoculation must be balanced by the enzymes added to obtain optimal results (Van Rensburg et al., 2013). This is a very promising process as it can reduce both capital cost and material cost, since the enzymes need not to be separately produced or bought in (Van Rensburg et al., 2013).

Different types of enzymes and temperatures can be used for liquefaction, which would influence the profitability of the plant. Therefore, in the next section different enzyme processes are discussed.

2.5 Enzyme process

There are two enzyme processes that can be used to liquefy starch, namely the conventional warm process and the cold process, see Figure 3. These two processes are discussed below while focussing on corn and triticale feedstock.

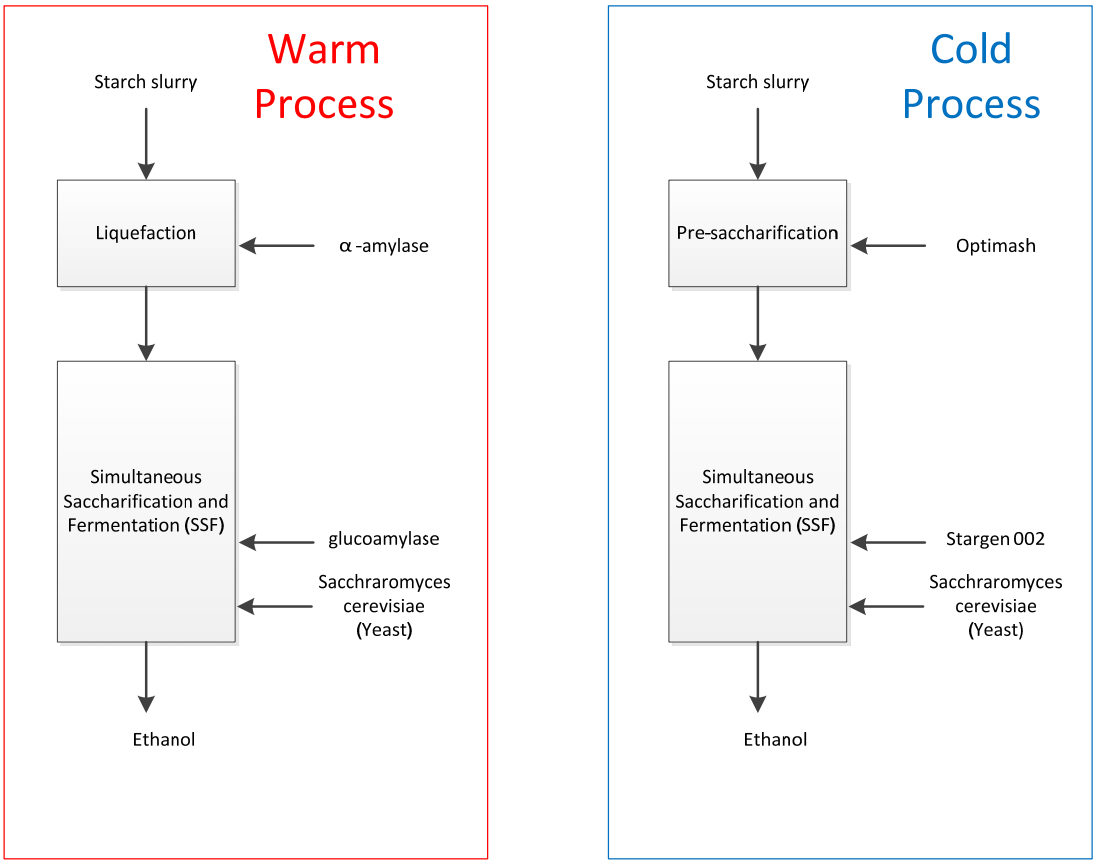

FIGURE 3: WARM AND COLD ENZYME PROCESS

2.5.1 Warm enzyme process

The warm process is conventionally used in the industry for the dry grinding process. It performs the first hydrolysis step of liquefaction at a temperature above the gelatinisation temperature of starch. Triticale can be liquefied (60°C) at lower temperatures than corn (90°C) or wheat (65°C) (Pejin et al., 2009). For liquefaction, grinded triticale is added to water and mixed to achieve a liquefaction mash. This mash is then heated to a temperature of 60°C and then α-amylase is added (Figure 3) (Pejin et al., 2009). Glucoamylase is only added in fermentation (Wang et al., 1998). The hydrolysis efficiency of the enzymes' starch to glucose conversion is close to 100% (Pejin et al., 2015) The ratio of triticale to water is 1:3. The pH of the mixture is adjusted to 5.4-5.5 and the resident time in the reactor is 60 minutes.

Additionally adding urea, calcium and magnesium also improves the ethanol yield of the process (Pejin et al., 2015; Wang et al., 1997). According to Vučurović and Pejin (2007) the best cultivar of triticale in Eastern Europe is Jutro. The best triticale cultivars in SA for ethanol yield are D1, H1 and D2 in the Swartland region and H1 and G2 in the Overberg region (Tsupko, 2009). The best triticale variety may vary with region and thus only triticale varieties grown in SA can be used in the models.

According to research in Eastern Europe almost no technical enzymes need to be added in the case of triticale because the technical enzymes only cause a slight increase in ethanol yield (36.26 to 38.5 g bioethanol/100 g dm) in the case of NST 21/06 and even less for the other varieties (Pejin et al., 2009). South African triticale varities are not bred for native enzymes as these grain germinate very easily which is undesirable (Willem Botes, 2014). Futhermore it should also be noted that the viscosity of the triticale mixture is low and thus viscocity is not a problem in the process (Pejin et al., 2009).

The warm process used for triticale is quite different from those conventionally used for corn in terms of temperature. A description of the warm process for corn follows. The corn mash is sent to a jet cooker, which ruptures the starch molecules. The jet cooker is operated at temperatures above 100°C. The corn mash is then usually cooled down to 90°C with a residence time of 30 minutes in the liquefaction reactor (Bothast and Schlicher, 2005). Thermo stable α-amylase is added during the warm process (liquefaction), but glucoamylase is only added in the fermentation reactor (Kwiatkowski et al., 2006).

To summarise, triticale can be liquefied at lower temperatures than corn. This can be attributed to the different gelatinisation temperatures of each. The lower temperature contributes to minimizing the cost of the warm process.

2.5.2 Cold enzyme process

In the cold enzyme process, the hydrolysis is performed at temperatures below the gelatinisation temperature. Operational cost decreases due to energy savings as a result of the lower operational temperature. This renders the process more energy efficient. The process is operated at temperature of 30°C with a resident time of 92 hours. The solid loading of the process is 30% (Li et al., 2012). Urea is added to lower the gelatinisation temperature, as it breaks the intermolecular bonding of starch molecules (McGrane et al., 2004). The enzyme used in the cold process is Stargen. The cold enzyme (Stargen 002) requires a pre-saccharification step to improve the hydrolysis efficiency. Pre-saccharification is performed, with an amylase product such as Optimash, at a temperature of 57ºC with a residence time of 120 minutes (Tsupko, 2009). The hydrolysis efficiency of Stargen 002 is 96% (Tsupko, 2009). This value of 96% is lower than the 98% achieved for the warm process. The cold enzymes are less efficient at converting starch to ethanol as can be seen through the

larger amount of resistant starch being left at the end of the fermentation.

Resistant starch is the unhydrolysed starch left over at the end of fermentation (Sharma et al., 2009). The resistant starch decreases by 11-43% for the cold enzymes and 55-74% for the warm enzymes (Sharma et al., 2009). This indicates the warm process have more fermentable starch (Sharma et al., 2009). Thus the cold process is less efficient at converting starch to glucose than the warm process.

In the cold process, the starch liquefaction present in the warm process is replaced with a pre-saccharification step, which for wheat is performed at 60°C for 30 minutes. The enzymes that are added for pre-saccharification are pulullanase and glucoamylase. These enzymes hydrolyse dextrins and this causes glucose and the other six carbon sugars to ferment more easily (Gago et al., 2013). Pre-saccharification can also reduce the viscosity of the mash for wheat straw (Paulová et al., 2014). According to literature, pre-activation of triticale is done at a temperature of 50°C and has a resident time of 30 minutes (Li et al., 2012). This temperature is 5°C lower than the gelatinisation temperature of triticale (Li et al., 2012). In the case of triticale, only urea was added to this process (Li et al., 2012).

Stargen used in the cold process is an enzyme blend of acid α-amylase and glucoamylase that is produced from *Aspergillus kawachii* and *Aspergillus niger.* It is produced by the company Genencor, which is part of the Dupont-Danisco group (Li et al., 2012). The glucoamylase drills holes in the starch that allow the α-amylase access to hydrolyse the starch from within. It is desirable that the starch has a higher concentration of amylopectin as it is easier hydrolyse than amylose by Stargen (Adams et al., 2012).

There are different types of process configurations depending on the by-products and feedstock. Different configurations are discussed in the next section.

2.6 Type of Process

There are different types of processes that can be used to produce bioethanol from triticale. Two of these are described below. They are the conventional dry grinding process and the pre-fractionation dry grinding process.

2.6.1 Conventional Dry Grinding Process

The conventional dry grinding process is used extensively in the USA for grain-ethanol production (Sharma et al., 2009; P. Wang et al., 2007). It is usually done in combination with the warm enzyme process but can also be done with the cold enzyme process (P. Wang et al., 2007). POET commercialized the cold process in 2004 and have 24 biorefineries in the USA (POET, 2015). Figure 4 shows the flowsheet for a conventional dry grinding facility with the warm enzyme process.

Grain is fed to a hammer mill where it is finely ground for easier hydrolysis (ICM Inc., 2012). It is then fed to the cook/slurry tanks to be mixed with water (ICM Inc., 2012). In the liquefaction tanks α-amylase is added and the starch is hydrolysed to glucose (ICM Inc., 2012). In the ethanol fermentation section, glucoamylase is added to break dextrins up to simple sugars. Yeast is also added to ferment glucose to ethanol and CO_2 (ICM Inc., 2012). A temperature of 35°C is used for the fermentation of triticale grain (Balcerek and Pielech-Przybylska, 2013). The residence time in the case of triticale that the mash spends in fermentation reactors is 48 hours (Wang et al., 1997). Other authors suggest a fermentation time of 50-60 hours for grain (Bothast and Schlicher, 2005; ICM Inc., 2012; Lin et al., 2011a). The liquid mixture from the ethanol fermentation is fed into distillation columns. The columns separate ethanol from the water and solids. The volume of ethanol present after distillation is 95%. (ICM Inc., 2012) The ethanol is further purified using molecular sieves to a purity of 99% (Amigun et al., 2012). Molecular sieves consist of tanks with specialised molecular sieve beads that absorb water but not ethanol. Denaturant is added next to make the ethanol unfit for human consumption. Finally, the ethanol can be stored, and is ready for sale (ICM Inc., 2012).

The solids and water (stillage) separated from the ethanol in the first distillation column are fed into a centrifuge. The centrifuge separates the solids from the liquids. The liquid steam contains 5-10% solids and is called thin stillage (ICM Inc., 2012). The solids' moisture content is 65 wt% and is called Wet Distillers Grains (WDG) (U.S. Grain council, 2012). The water in the liquid stream is subsequently evaporated to decrease the moisture content to 60 wt%. Other sources refer to concentrations of solids after evaporation as between 25-50%(ICM Inc., 2012) (ICM Inc., 2012). In the syrup tanks, the WDG and evaporated liquid stream (syrup) are mixed together. This mixture is further dried in a rotary drum to achieve a moisture content of 10-12 wt%. The resulting product is then called Dried Distillers Grains and Solubles (DDGS), which is classified as a by-product (U.S. Grain council, 2012).

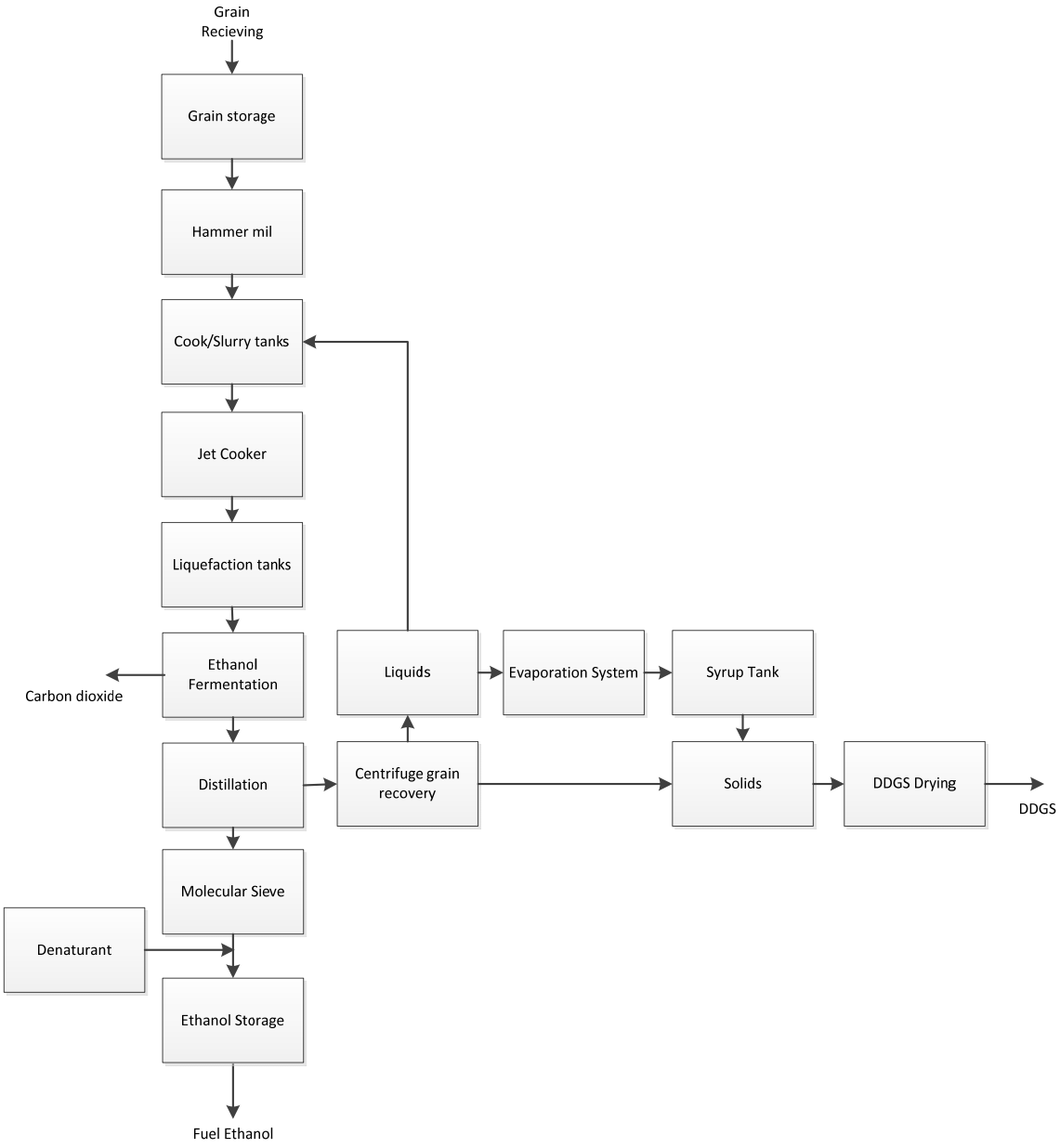

FIGURE 4: BLOCK FLOW DIAGRAM OF CONVENTIONAL DRY GRINDING

REDRAWN FROM: (Amigun et al., 2012)

2.6.2 Pre-fractionation Dry Grinding Process

The pre-fractionation dry grinding process is similar to the conventional dry grinding process except for that the bran is removed from the grain prior to the grinding step (Amigun et al., 2012, 2011). The bran is then used in the boiler for producing steam for the process. The protein content of DDGS increases as a result of pre-fractionation, due to the removal of the bran that is high in fibre (Amigun et al., 2012, 2011). Therefore, less DDGS is produced but with a higher protein content. The bran used to produce the steam also lowers the operational cost (Amigun et al., 2012, 2011). Furthermore, according to Amigun (2012), the pre-fractionated DDGS can be priced as soybean meal, which is priced higher than conventional DDGS. See Figure 5 for a flow diagram of the process.

According to Nkomba (2015) the dry pre-fractionation process gives a low ethanol yield if the bran is not enzymatically treated. This is due to the protein and starch contents of the bran, which provide essential nutrients required for yeast growth and starch fermentation. Therefore, it is recommended that the bran be treated for six hours at a temperature of 30ºC with amylases (Stargen 002) and proteases (Fermgen) to hydrolyse the starch and protein from the bran (Nkomba, 2015). The resulting hydrolysate is subsequently fed to the fermenter as make-up water.

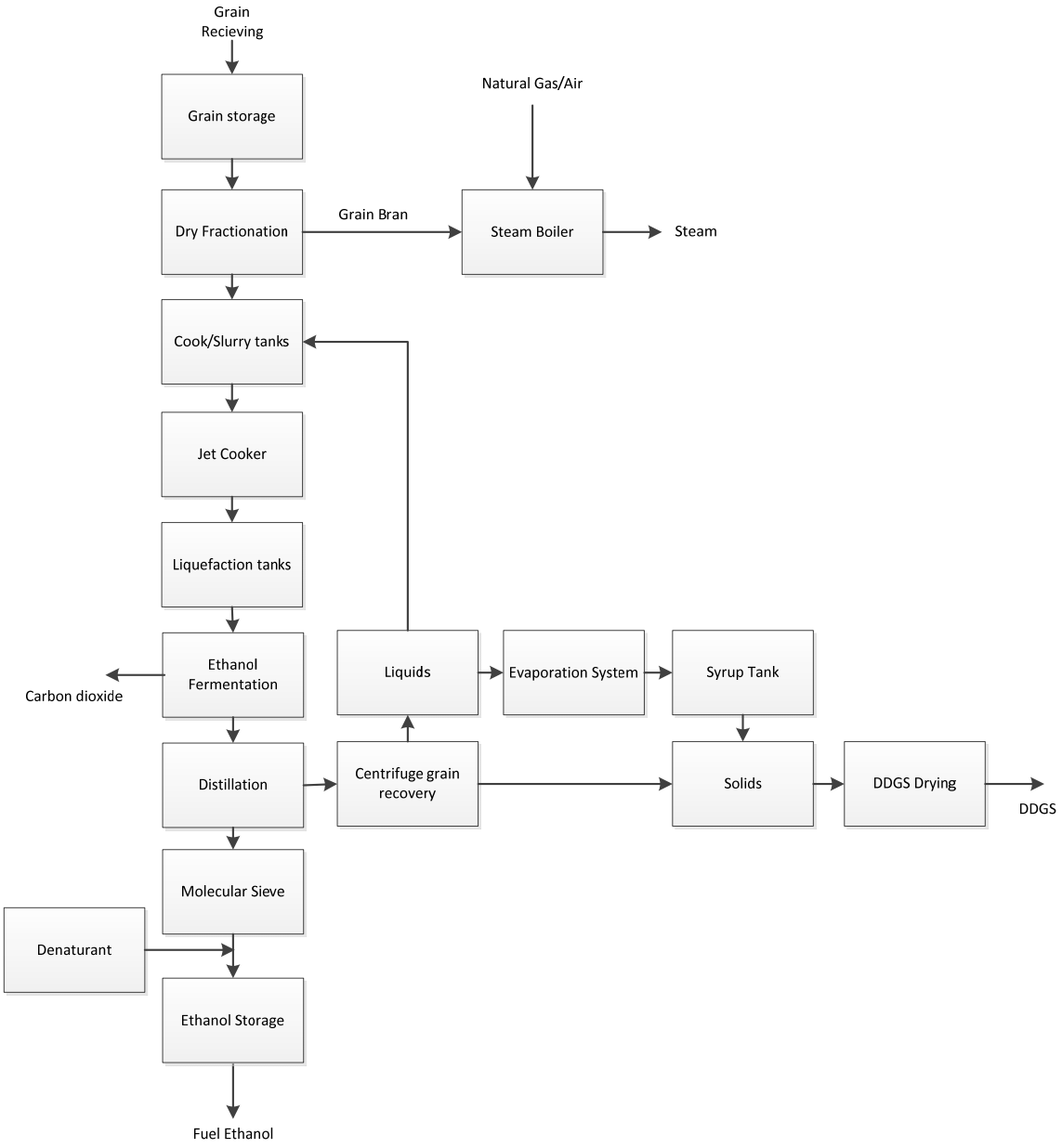

FIGURE 5: BLOCK FLOW DIAGRAM OF DRY PRE-FRACTIONATION GRINDING PROCESS

REDRAWN FROM: (Amigun et al., 2012)

2.7 Economic Feasibility

The economic feasibility of the different processes needs to be evaluated. The following main aspects affect the process/plant profitability, namely the capital cost, the operational (manufacturing) cost, the product selling price (bioethanol selling price) and the selling price of the by-products. The Internal Rate of Return (IRR) of the process determines if the investment will be profitable for investors and banks and is therefore considered in this section.

2.7.1 Capital Cost

According to literature, the capital cost of a starch bioethanol plant in developed countries is \$1.17-2.13/gal (R2.29-8.46/ℓ), but the projected capital cost in SA is \$6.94 /gal (R23.32/ℓ) (Kwiatkowski et al., 2006; McAloon et al., 2000; Tait, 2014). The higher cost in SA is probably due to high shipping costs and import taxes as well as over quotes in equipment capital cost.

The scale up of the capital cost from the capacity of an existing plant to obtain an estimate for the capacity of the considered plant is calculated in equation 1. Here C_1 is the cost of the existing plant at capacity 1 (Q_1) while C_2 is the cost of the considered plant at capacity 2 (Q_2). The scale up or down factor is X, which is 0.6 for a plant size of between 80 and 160 million L ethanol per year (Amigun et al., 2012).

$$C_2 = C_1 \left(\frac{Q_2}{Q_1}\right)^X \quad [1]$$

The calculated cost of the plant being considered should also be adjusted for the inflation experience over the scale-up time period. The adjustment /correction is done in dollar and hence all the capital costs should also be converted to dollars if necessary (Turton, 2009). The required equation for the adjustment is as follows:

$$C_{2,corrected} = C_2 \left(\frac{I_2}{I_1}\right) \quad [2]$$

I_2 is the Chemical Engineering's Plant Cost Index (CEPCI) value in the year that the cost of the plant needs to be determined for (in the present/future), while I_1 is the historical CEPCI value for the year the plant cost was originally determined (in the past).

2.7.2 Manufacturing Cost

The production (operational) cost consists mostly of the cost of buying triticale grain. The cost of feedstock is therefore a very important factor in the determination of the profitability of the plant (Amigun et al., 2012). The cost of triticale is approximated as the price of B4 wheat on the South African Futures Exchange (SAFEX) index (see

Figure 6) for last ±6 years (SAFEX-JSE, 2015). The SAFEX price was corrected for the transport differential (R570/tonne) between the WC and SAFEX (located in Randfontein, Gauteng), to obtain an assumed price for triticale grains in the WC (Coetzee, 2015) Yeast is expensive (R150/kg), but is used in small amounts (0.5g/ℓ). Electricity, water and enzyme prices can also affect the cost of manufacturing (Amigun et al., 2011).

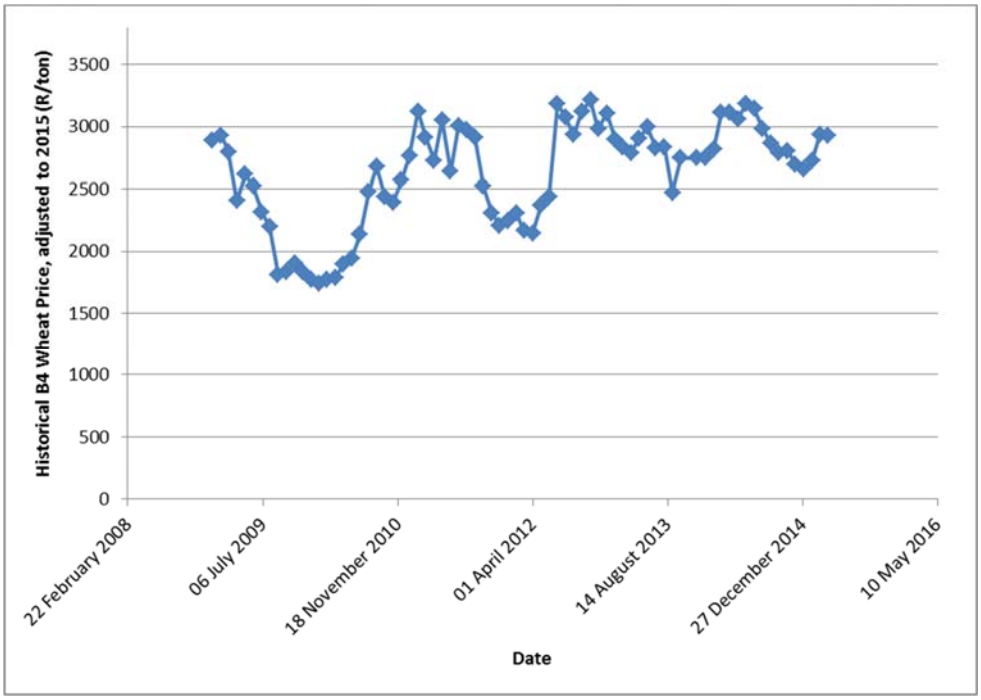

FIGURE 6: HISTORICAL B4 WHEAT PRICES FROM JANUARY 2009 - APRIL 2015

2.7.3 Ethanol Selling Price and Proposed Biofuels Subsidy in South Africa

The ethanol selling price is compared to the basic fuel price (BFP), based on the "blending value" of the ethanol. For 2% blending the ethanol selling price can be assumed to be the same as the BFP (Department of Energy, 2014a), as the blending value is 1; see equation 3.

$$Ethanol\ Price\ at\ refinery = BFP\ x\ Blending\ value \quad [3]$$

As ethanol is the primary product, the ethanol selling price has a major effect on the plant's profitability. The minimum ethanol selling price (MESP) is the price ethanol needs to be sold at to achieve a Return on Assets (ROA) of 15%; see equation 4 (Department of Energy, 2014a).

$$ROA = \frac{Earnings\ before\ tax\ and\ interest}{Total\ Assets} \quad [4]$$

The ROA's "Earnings before tax and interest" is the actual cashflow from operations, measured as EBITDA (Earnings before Interest, Tax, Depreciation and Amortisation) in the Income Statement for a particular time period. The total assets is the Asset Value from the Balance sheet for the same time period. If this value is less than 15%, then a subsidy value is added such that EBITDA plus a subsidy divided by asset value equals 15% (Department of Energy, 2014a).

The 15% ROA, which is specified by the government, is used to determine the subsidy needed for project viability. The government defines economically viability as 15% ROA. If the MESP is higher than the "*BFP x blending value*" a subsidy from the government is needed to render the process profitable. If the MESP is lower than "BFP x blending value", no subsidy is needed as the 15% ROA is achieved. The MESP equation relationship is as seen in equation 5 below (Tait, 2014):

$$MESP = subsidy + BFP\ x\ Blending\ value \quad [5]$$

ROA calculation needs to be done for a reference plant (Department of Energy, 2014a). Three different reference plants are used, namely for a sorghum (starch feedstock), a sugarcane (sucrose feedstock) and a soya oil (biodiesel) biofuel plant (Department of Energy, 2014a). The commercial plant's subsidy will be based on the subsidy calculated for the closest reference plant.

BFP prices are shown from December 1995 to April 2015; see Figure 7 for a range of BFP prices. The ethanol yield from triticale varies between 309kg/tonne triticale (Amigun et al., 2011), 377kg/tonne triticale (Tsupko, 2009) and 470kg/tonne triticale (Amigun et al., 2012). This yield depends on the variety of triticale used.

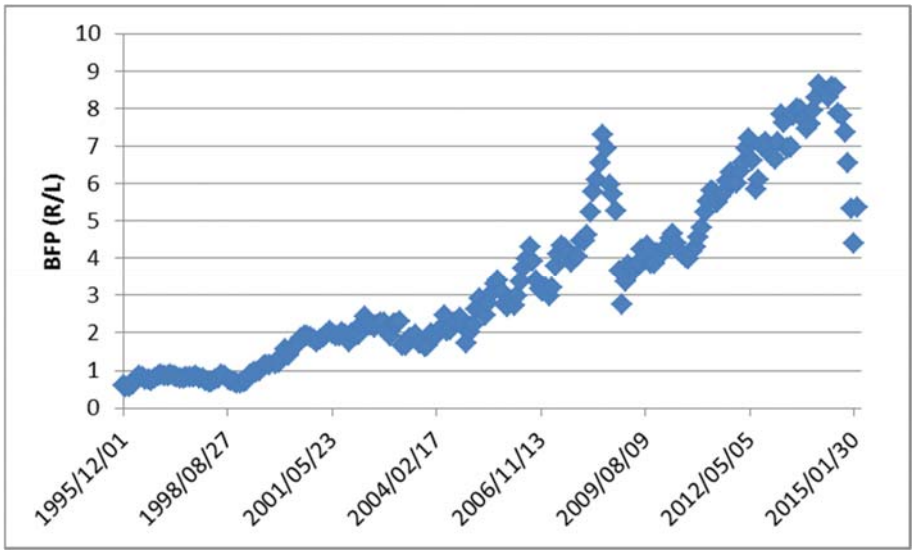

FIGURE 7: HISTORICAL BFP FROM DECEMBER 1995 - APRIL 2015

2.7.4 Selling Price of by-products

The by-products of the bioethanol process are DDGS and CO_2. In the conventional dry grinding process the DDGS contains 35-45% protein (Best et al., 2005; Kwiatkowski et al., 2006), whereas in the modified pre-fractionation process the protein content is 65% (Best et al., 2005). Therefore, DDGS from the pre-fractionation process can be used as a feedstock for non-rumen animals like swine and poultry (Singh et al., 1999). The price of the higher protein content DDGS can be approximated as the price of soya oil cake, which is a high value product (Ten Cate, 2015a). Soya oil cake is imported into SA and therefore there is demand for it. This can lead to an increase in the profitability of the process (Amigun et al., 2011). For every tonne triticale, 333 kg DDGS as well as 333 kg CO_2 are produced in the conventional dry grinding process (Amigun et al., 2011).

2.7.5 The Internal Rate of Return (IRR)

The IRR of the process can be used to determine if the investment will be attractive for investors. Furthermore, it can also be used by banks to determine if they will finance the capital of the plant as well as if the risk associated with their capital expenditure is acceptable. The IRR is the interest rate needed to achieve a Net Present Value (NPV) of zero. See the IRR formula in equation 6 below (Turton, 2009):

$$0 = NPV = \sum_{n=1}^{N} Cashflow_n \times (1+i)^{-n} \quad [6]$$

Here *N* is the plant's lifespan (combined construction and operational period), *n* is the applicable project year and *i* is the interest rate. The interest rate (*i*) for the IRR

calculation is the IRR itself. The IRR is calculated iteratively.

The NPV is the sum of all discounted cash flows. It is therefore the present value of all future cash flows of the plant. See the NPV formula in equation 7 below:

$$NPV = \sum_{n=1}^{N} Cashflow_n \times (1+i)^{-n} \quad [7]$$

2.8 Energy and energy fuels alternatives

The conversion of grains into ethanol requires significant amounts of process energy in the form of low pressure steam and electricity (Amigun et al., 2011 and 2012). Onsite production of process steam can be performed with a low boiler (Process Heat = PH) in combination with the buy-in of electricity from external suppliers such as the national grid or with a CHP plant. A Combined Heat and Power (CHP) plant generates both low pressure steam and electricity for the process plant. Electricity production by a CHP plant, sized according to the process steam demand, is usually in excess of the process electricity demand. Surplus electricity is therefore provided that can be sold to nearby users (Kaliyan et al., 2011). A CHP plant consists of a high pressure boiler and a series of turbines. Figure 8 shows the configuration of a CHP plant (Nsaful et al., 2013). Usually the pressure at the last turbine's outlet should be equal to that of the amount of process steam needed for the process.

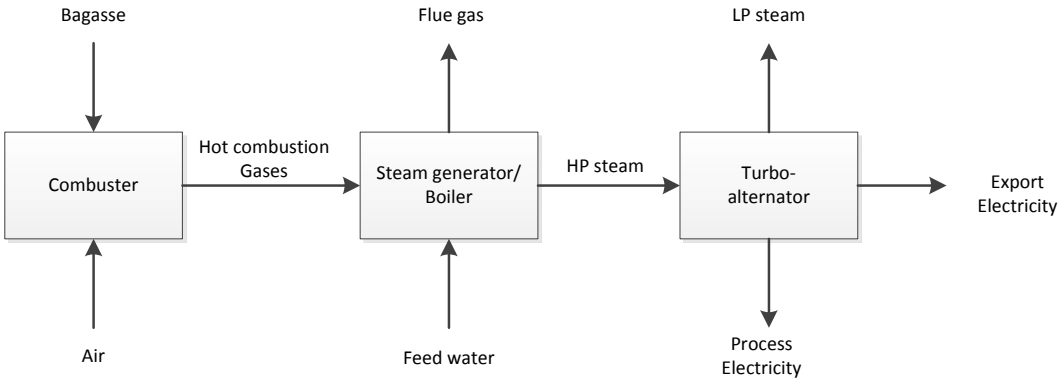

FIGURE 8: CHP PLANT CONFIGURATION

REDRAWN AND MODIFIED FROM: (Nsaful et al., 2013)

According to literature, fuelling a CHP plant with coal to supply steam and electricity to a grain-ethanol plant will cause a significant environmental burden (M. Wang et al., 2007). This is due to the emissions associated with coal combustion. As a result, only a minor reduction of approximately 2% in the GHG emissions, over the life cycle of corn production, ethanol production and ethanol use (compared to gasoline based on crude oil) will be obtained (cradle to grave approach) (Wang et al., 2007). The minor GHG benefit is essentially as a result of avoided electricity transport losses, as

electricity is produced and consumed onsite (Wang et al., 2007). Additionally, it is estimated that if a CHP plant uses biomass, the reduction in GHG emissions can be as large as 101.3% over the whole corn-ethanol life cycle as the fuel source is renewable and carbon neutral (Kaliyan et al., 2011). For sorghum, the ethanol GHG emissions reductions are 71% and for sugar they are 70% when the life cycle of the grain-ethanol is compared to the gasoline life cycle (Cai et al., 2013).

Steam can also be produced by using different fuels such as coal or biomass for steam generation. Using biomass as an energy source for steam generation lowers the ethanol life cycle GHG emissions compared to gasoline life cycle emissions by 52%. Using coal as energy source increases the corn ethanol GHG life cycle emissions (compared to gasoline life cycle emissions) by 3% and therefore ethanol produces more GHG emissions than gasoline does (M. Wang et al., 2007). Therefore, to render ethanol a more environmentally friendly fuel than gasoline, biomass should be used as energy source for steam generation.

2.9 Similar Studies

The closest study to this current study would be Amigun et al. (2011; 2012). They are both done for the WC and use at triticale as feedstock. In this study other grains like wheat and barley (Amigun et al., 2012). This study also investigated different process configurations like the removal of bran from the conventional dry grinding process (pre-fractionation process) as well as the removal of the bran in combination with fermentation of the bran to produce cellulosic bioethanol (Amigun et al., 2012). The process was not modelled in a process simulator like Aspen but fermentation data was use as well as other estimates for process energy from literature sources for corn (Amigun et al., 2012). Some biomass namely bran is considered for energy production (partial boiler feedstock) (Amigun et al., 2012). The subsidy mechanism investigated was 100% fuel levy exemption (±R1.5/ℓ). The plant was assumed to use 200 000/ton grain feedstock per year (±78 million ℓ ethanol/y) (Amigun et al., 2012). Amigun (2011) did a Monte Carlo simulation to determine probability of plant profitability. It was found that the pre-fractionation process with the subsidy mechanism assumed, has a 98% chance of economic success (Amigun et al., 2011).

Amigun et al. (2011;2012) study was influenced by Best et al. (2005). Best et al. (2005) was done for corn in the Georgia in USA. They investigated the warm dry grinding process and pre-fractionated process for corn (Best et al., 2005). They also did a Monte Carlo simulation and found grain and ethanol price have the largest effect on the profitability of the plant (Best et al., 2005). They recommended also the pre-fractionated process for corn bioethanol production (Best et al., 2005).

Wheat as possible feedstock in the WC was investigated in a study by Richardson et al (2007). The objective of the study was to quantify risk and the viability of ethanol production in the WC through a Monte Carlo simulation (Richardson et al., 2007).

The study used USA prices and literature results for the Monte Carlo simulation (Richardson et al., 2007). A 103 million ℓ ethanol plant/y conventional dry grinding process was assumed for this study (Richardson et al., 2007). It was found that the plant would not be profitable. Therefore the government involvement through subsidy and other policy assistance will be needed to render bioethanol production in the WC economically viable (Richardson et al., 2007).

Kwiatkowski et al. (2006) simulated a conventional dry grinding process for corn in USA with the process simulator SuperPro Designer. This model was used and upgraded by Lin et al. (2011) to a dry pre-fractionation process. This study also assumed the plant will produced 136 million ℓ ethanol plant/y (Kwiatkowski et al., 2006; Lin et al., 2011b). The study found that pre-fractionation reduces energy and water consumption and increase the profitability of the plant (Lin et al., 2011b).

Nkomba (2015) investigated the warm, cold and pre-fractionated process for sorghum. The cold and warm process had similar ethanol yields (Nkomba, 2015). The warm pre-fractionation process preformed worse than the cold pre-fractionation process due to yeast nutrient deficiencies Nkomba, 2015).

Pieters (2016) investigated the warm and the cold process for corn. The study used experimental data from the study to create an Aspen model (Pieters, 2016). The Aspen model is the used to create an economic model to evaluate what is the highest price Stargen 002 can be brought at for the process to be still profitable. The maximum allowable Stargen 002 price was R498/kg (Pieters, 2016). According to Pieters (2016) the cold process will be more profitable than the warm process.

Ou et al. (2014) investigated the effect of a CHP plant (using corn stover) and cellulosic ethanol for corn compared to the conventional warm dry grinding process. The models was created using ChemCAD (Ou et al., 2014). This study also assumed the plant will produced 363 million ℓ ethanol plant/y (Ou et al., 2014). It has been reported from this study that a corn ethanol plant using a CHP plant operating on corn stover will achieve the lowest minimum ethanol selling price when compared to process using process heat (PH) (Ou et al., 2014). Thus using a CHP plant would be more profitable than using PH for a corn ethanol plant (Ou et al., 2014).

2.10 Summary

Bioethanol has both advantages and disadvantages above gasoline. It can be produced by different types of feedstocks but this study focus on starch-containing feedstock. Of all the starch-containing feedstocks, triticale shows the most promise in the WC. There are different types of processes and combinations of processes for hydrolysis and fermentation, namely SHF, SSF and CBP. There are two enzyme processes that can be used to liquefy starch, namely the conventional warm process and the cold process. The conventional warm dry grinding process can be modified

to remove bran from the grain. This will increase the protein content of DDGS, which can then be fed to non-rumen animals and therefore can be sold at a higher price. The BFP and triticale price show significant variation in price ranges. A 15% ROA subsidy mechanism is mandated by government to alleviate price fluctuations in product and feedstock prices. Bioethanol production can also significantly reduce GHG emissions if biomass is used as fuel source for energy production. Different similar studies were investigated but none was found that combine the warm, cold and pre-fractionation processes in an economic study for any grain. Furthermore, none of the process investigated a CHP plant in combination to the triticale-ethanol process. Therefore this study is unique as it is an economic study done for triticale grain in WC that considers the following process configurations namely warm, cold and pre-fractionation processes as well as the effect of a CHP plant will have on the economics of the process under the current 15% ROA subsidy mechanism.

The methodology used for the project follows in the next section. The literature study was used as basis for the methodology.

3 RESEARCH DESIGN AND METHODOLOGY

Process modelling of alternative process scenarios was performed, to compare energy balances and the economic viability of ethanol production from triticale. The descriptions of various steps in these process models were obtained mostly from published literature of experimental measurements, either for triticale itself or for grain-ethanol processes very similar to triticale. The models proposed in this section represent an attempt to answer the key questions listed in section 1.2. In this section the methodology for the process models will be discussed as well as the assumption and legislation pertinent to the process models. SA legislation stipulates the ethanol selling price as well as the subsidy mechanism needs to be use for bioethanol production. The methodology and assumptions needed for the economic models will also be considered. The methodology for the sensitivity and historical data analyses will be discussed in the economic section. Lastly, the methodology and assumptions for the CO_2 balance are discussed in the environmental section of this chapter.

3.1 Process Models

In this section the descriptions of the different Aspen models will be presented as well as the process assumptions needed to create these models.

3.1.1 Process description

Different models were built in Aspen *version 8.3* to evaluate which process provides the best results and is the most economically feasible. The models are as follows:

- The Conventional Dry Grinding Process with warm enzyme process (Model 1);
- The Dry Grinding Process with cold enzyme process (Model 2);
- The Pre-fractionation Dry Grinding Process with warm enzyme process (Model 3);
- The Pre-fractionation Dry Grinding Process with cold enzyme process (Model 4).

In the following sections the four process descriptions and associated flow sheets are presented.

3.1.1.1 *The Conventional Warm Dry Grinding Process (Model 1)*

The base case model is the “Warm” Dry Grinding Process, which is the conventional process for industrial production of ethanol from starch grains. The process is illustrated in Figure 9.

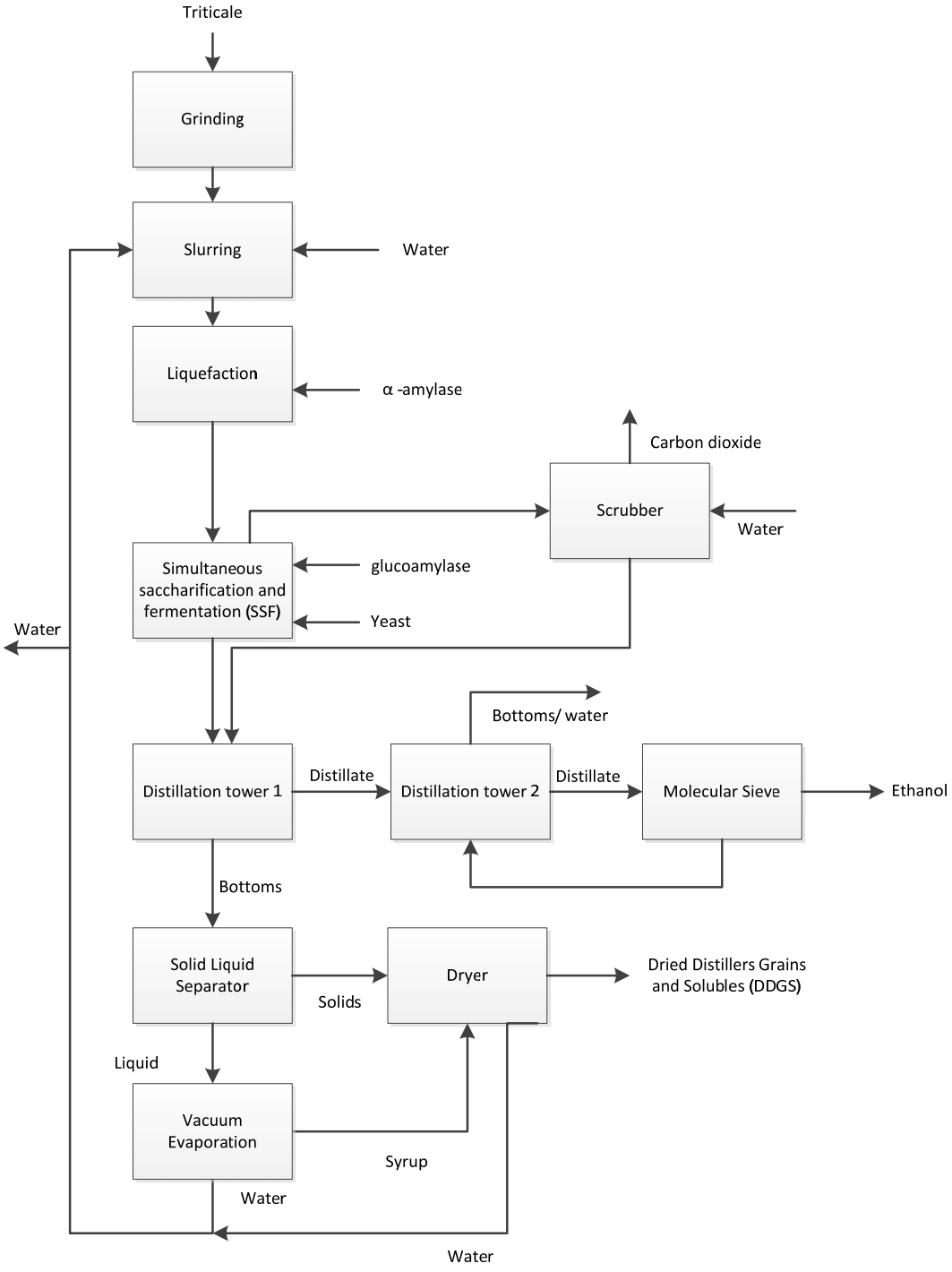

FIGURE 9: BLOCK FLOW DIAGRAM OF CONVENTIONAL DRY GRINDING PROCESS WITH WARM ENZYME PROCESS (MODEL 1).

Triticale is ground and mixed with water to obtain a solids loading of 30% (Vučurović and Pejin, 2007). The next stage is liquefaction, where the mixture is heated up to 60°C for approximately 60 minutes. Pejin et al. (2009) points out that α-amylase (see

"Liquefaction" reactor - Figure 9) and glucoamylase (see "SSF" reactor - Figure 9) enzymes are added for the warm process. The conversion of starch to glucose is assumed to be 98% of the theoretical yield, based on starch content of the raw material (Pejin et al., 2015). The mixture is then sent to the Simultaneous Saccharification and Fermentation (SSF) reactors, where yeast and glucoamylase enzymes are added and the operating temperature is 35°C (Balcerek and Pielech-Przybylska, 2013). The residence time of the mash in the SSF reactors is 48 hours (Wang et al., 1997). The ethanol yield of the SSF reactors is 95%, based on sugars released by enzymatic hydrolysis.

Carbon dioxide (CO_2), produced by the fermentation of glucose to ethanol, is removed from the SSF reactors in a vapour stream outlet. This stream also contains some ethanol that requires removal in a scrubber (Petersen et al., 2014). The scrubber recovers 99% of ethanol vapour and the resultant liquid is sent back to the first distillation column (Beer column). Two knock-out drums remove the excess CO_2 from the liquid stream before distillation. The first knock-out drum is operated at 86°C and 0.85 atmospheres and largely vaporises CO_2, some water and ethanol (Petersen et al., 2014). The second knock-out drum is operated at 70°C and 1 atmosphere and condenses most of the ethanol vapours, while most of the CO_2 remains in the vapour phase (Petersen et al., 2014). The two knock-out drums therefore remove 99% of the CO_2 in the liquid stream from the scrubber, while the residual liquid is sent back to distillation for ethanol recovery.

The fermented liquid mixture from the SSF reactors is sent to the distillation columns, in combination with the CO_2–free liquid from the vapour scrubber. The beer column removes solids and reduces the amount of water, prior to feeding it to the second column (Rectifier column). The mass fraction of ethanol in the distillate from the Beer column is 35% (Petersen et al., 2014). The Beer column contains only a re-boiler and not a condenser, as the distillate is sent to the Rectifier column in a vapour phase (Dias et al., 2011). The distillate from the Beer column is sent to the Rectifier column in vapour form, where the mass fraction of ethanol in the final product stream is 91% wt. (Amigun et al., 2012). The feed is in a vapour form to reduce the energy consumption of the plant. The ethanol condensate from the Rectifier column is further dehydrated by a molecular sieve system to achieve a final ethanol concentration of 99.9% Amigun et al., 2012). Denaturant is subsequently added to the purified ethanol to make it unfit for human consumption, prior to storage and sale (ICM Inc., 2012).

The solids (bottoms) from the Beer column are sent to a solid-liquid separator. The moisture content of the resulting solid product is 65 wt%, which is called Wet Distillers Grains (WDG) (U.S. Grain council, 2012). The water in the liquid stream from the solid-liquid separator is processed through the use of vacuum evaporation, to decrease the moisture content to 60 wt%. The resulting evaporated liquid stream (syrup) and the WDG are mixed and further dried in a rotary drum to obtain a

moisture content of 12 wt% (U.S. Grain council, 2012). This product is then called Dried Distillers Grains and Solubles (DDGS), which is an animal feed by-product of the process (U.S. Grain council, 2012).

3.1.1.2 *The Cold Dry Grinding Process (Model 2)*

The cold process (Figure 10) differs from the warm process in that the enzyme cocktail used in mashing (such as Stargen 002 from Genencor) is capable of performing hydrolysis at a lower temperature (30ºC) than the warm process (60ºC) (Li et al., 2012; Pejin et al., 2009). The "cold" enzyme cocktail has a lower hydrolysis efficiency (96%), compared to that found in the warm enzymes (98%), but requires lower energy inputs for starch conversion to fermentable sugar (Tsupko, 2009). The cold enzymes do require a pre-saccharification step to improve the hydrolysis efficiency: Pre-saccharification is performed with an amylase product, like Optimash, at a temperature of 57ºC with a residence time of 120 minutes (Tsupko, 2009).

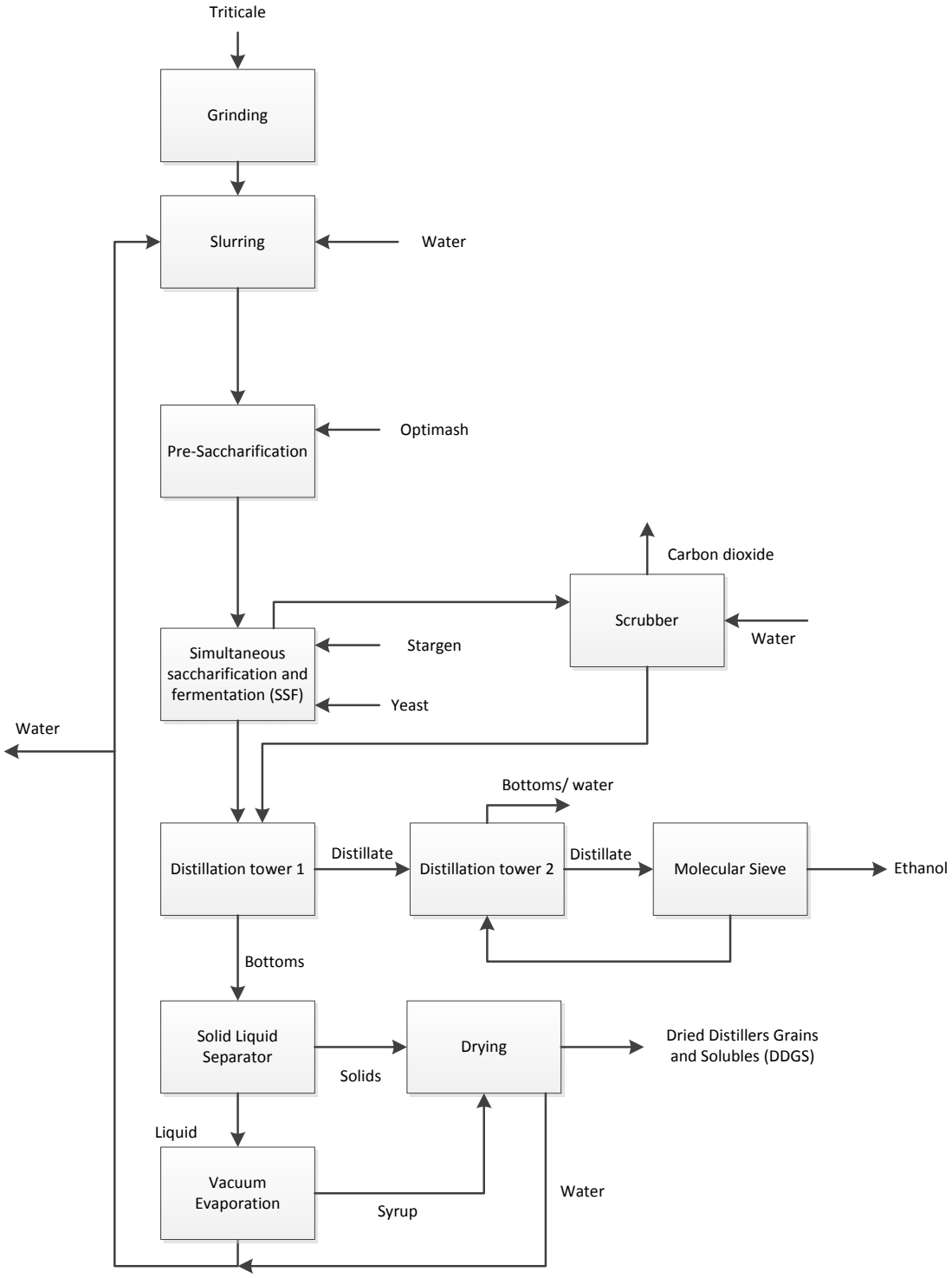

FIGURE 10: BLOCK FLOW DIAGRAM OF DRY GRINDING PROCESS WITH COLD ENZYME PROCESS. (MODEL 2)

3.1.1.3 *The Warm Pre-fractionation Dry Grinding Process (Model 3)*

The Pre-fractionation Dry Grinding Process (Figure 11) is similar to the conventional dry grinding process, but differs in that the bran is removed from the triticale grain prior to final grinding for hydrolysis-fermentation (Amigun et al., 2012). The bran separated from the triticale grain is enzymatically treated to hydrolyse the residual protein and starch contents. This occurs in the Bran Hydrolysis reactor, operated at a temperature of 30ºC and with a residence time of 6 hours (Nkomba, 2015). The liquid from the Bran Hydrolysis Reactor is separated from the solids by a solid-liquid separator and used as make-up water in the SSF reactor. This liquid product is combined with the starch-rich solids from Liquefaction to form a combined feed to the SSF reactor. This ensures that protein and residual starch in the bran are recovered as raw materials for fermentation (nutritional source for the yeast; additional carbon source for ethanol) (Nkomba, 2015). The spent bran, consisting mostly of fibrous lignocellulose, can be used as an energy source in the boiler, where process steam is produced.

The Pre-fractionation Dry Grinding Process significantly decreases the DDGS's fibre (bran) content and therefore lowers the mass yield of the DDGS. As a result, DDGS with a higher protein content, compared to that in the "Warm" and "Cold" Dry Grinding Process, is produced. Although a lower quantity of DDGS is produced, it has a significantly higher market value due to the higher protein content (Amigun et al., 2012, 2011).

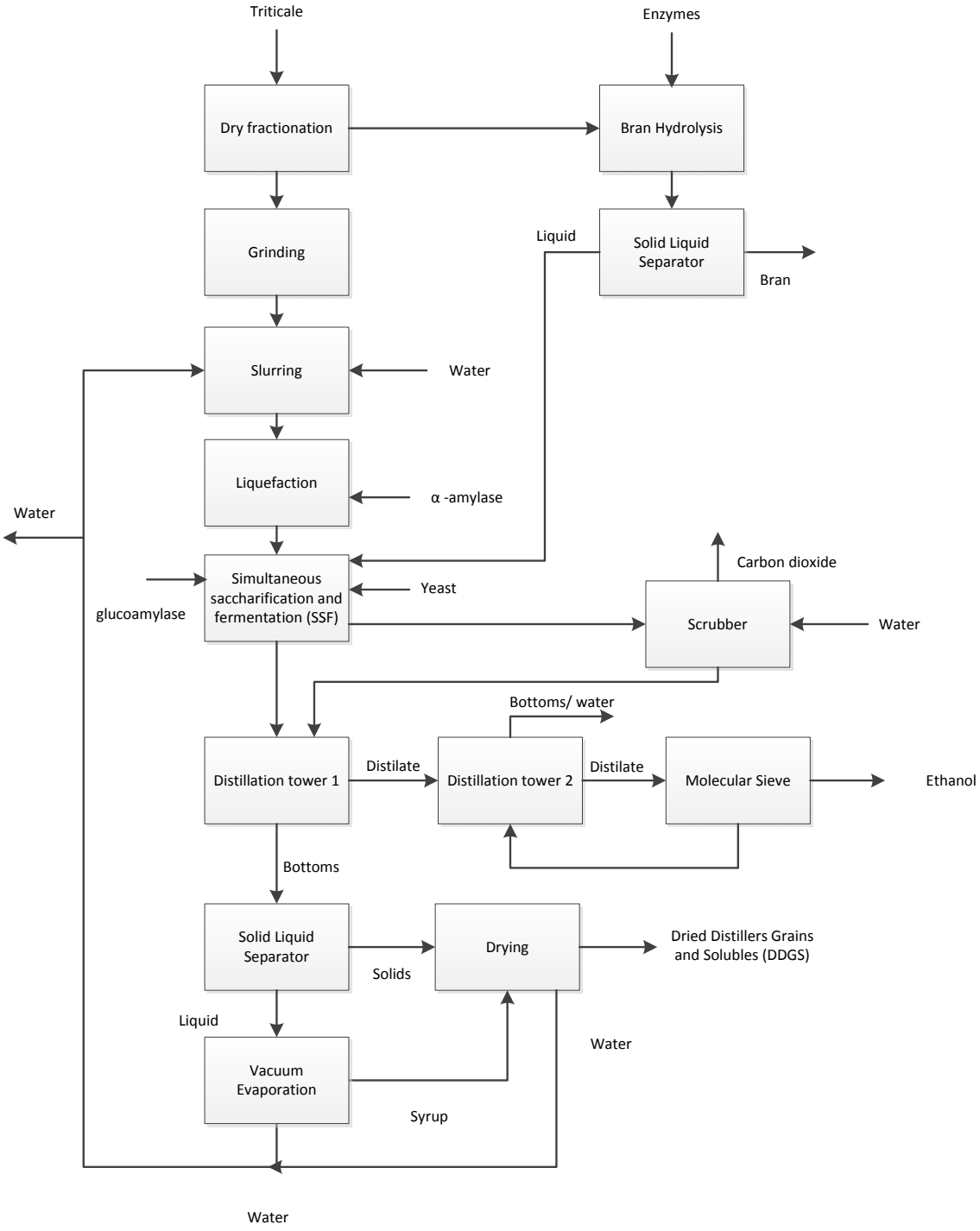

FIGURE 11: BLOCK FLOW DIAGRAM OF DRY PRE-FRACTIONATION GRINDING PROCESS WITH WARM ENZYME PROCESS (MODEL 3).

3.1.1.4 *The Cold Pre-fractionation Dry Grinding Process (Model 4)*

The Cold Pre-fractionation Process is the same as the warm Pre-fractionation Process except for the fact that the process makes use of cold process enzymes for hydrolysis. Therefore, this process is a combination of the pre-fractionation process and the cold process; see Figure 12 for the process diagram.

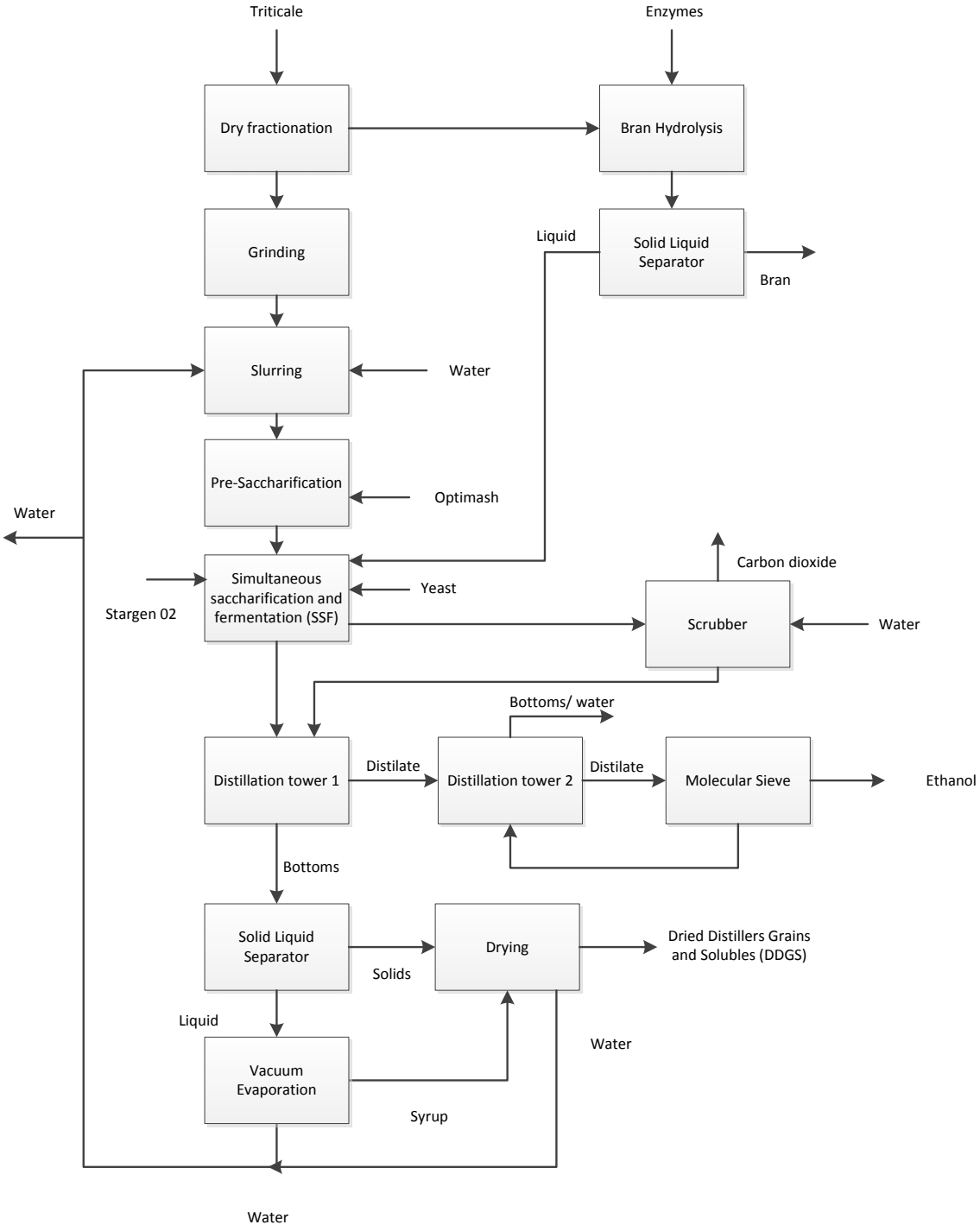

FIGURE 12: BLOCK FLOW DIAGRAM OF DRY PRE-FRACTIONATION GRINDING PROCESS WITH COLD ENZYME PROCESS (MODEL 4).

3.1.2 Process Assumptions

Certain assumptions are needed to model the processes. The process assumptions that are made for all Aspen models are listed below:

- The yield of ethanol from triticale is 435 ℓ/tonne.
- The process is operated at steady state.
- Perfect mixing is assumed for the process.
- The solid loading of the process is 30% (Vučurović and Pejin, 2007).
- 3% of all water used in the process and utilities is purged (Walas, 1990).
- A yeast inoculum mass percentage of 0.13% (yeast inoculum/triticale) is used in the process (Van Rensburg, 2014).
- The Stargen 002 dosage is 1.2kg/tonne triticale for the cold processes (Danisco US Inc, 2009).
- Triticale grain is assumed to consist only of starch, cellulose, xylan, water and protein for the Aspen models (Humbird et al., 2011).
- Triticale bran composition after pre-fractionation and before enzymatic hydrolysis is 20% starch, 17% cellulose, 13% xylan (García-Aparicio et al., 2011) and protein 17%.
- The calorific value (Cv) of coal is 7.5kWh/kg (Biomass Energy Centre, 2011).
- The calorific value (Cv) of biomass (30% wet) is 3.5kWh/kg (Biomass Energy Centre, 2011).
- The lower heating value (LHV) of coal is 22732 kJ/kg (Boundy et al., 2011).

Listed in the Table 2 below are the assumptions that differ in different models.

TABLE 2: PROCESS ASSUMPTION THAT DIFFER IN DIFFERENT MODELS

Processes	Warm Process (Model 1)	Cold Process (Model 2)	Warm Pre-Fractionation Process (Model 3)	Cold Pre-Fractionation Process (Model 4)
Ethanol yield	95%[12]	95%	95%	95%
Glucose yield	98%[13]	96%[14]	98%	96%
Operating temperature for liquefaction	60°C[15]	30°C[16]	60°C	30°C
Pre- saccharification step and temperature	No	Yes[17]	No	Yes
Operating temperature pre-saccharification	-	57°C[18]	-	57°C
Protein and starch recovery from bran	0%	0%	90%[19]	90%
Bran Removed	0%	0%	90%	90%

[12] From (Kasavi et al., 2012)
[13] From (Pejin et al., 2015)
[14] From (Tsupko, 2009)
[15] From (Pejin et al., 2009)
[16] From (Tsupko, 2009)
[17] From (Tsupko, 2009)
[18] From (Tsupko, 2009)
[19] From (García-Aparicio et al., 2011)

3.2 Economic Models

In the following section the methodology and assumptions made and legalisation needed for the economic models are given. The methodology for the sensitivity analysis to determine which factors influence the economic viability process the most is also given. Additionally the methodology for the historical data analysis to determine process profitability is given.

3.2.1 Methodology for Economic Models

Sixteen economic models were created to test different process configurations, fuels used for steam generation and the inclusion of a CHP plant with surplus electricity sales, and their impact on the profitability of the plant. In Figure 13 there is a breakdown of the economic models created for the project. The subsidy for each of the 16 models were calculated in two different ways, namely assuming straight line

depreciation or assuming no depreciation with asset values in real terms remaining constant over the life of the plant. The latter method will clearly incur larger subsidy support, which will positively influence the profitability of the plant. It should also be noted that all economic calculations are done in real terms, and nominal values are converted into real values. The economic calculations for the models are partly done in Aspen, from where the data is then extracted and incorporated into Microsoft Excel spreadsheets.

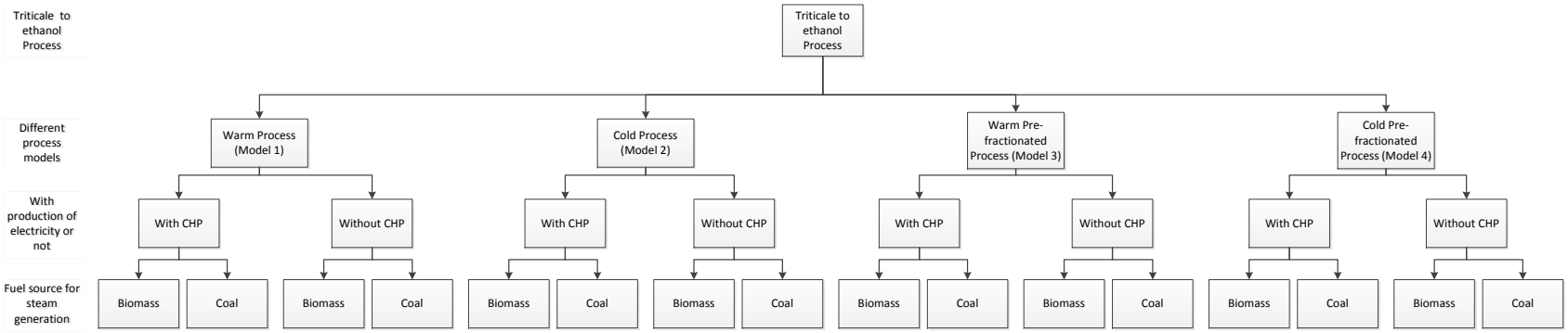

FIGURE 13: BREAK DOWN OF MODELS FOR ECONOMIC ANALYSIS

3.2.1.1 *Process variations*

The four process simulations/models built in Aspen and considered for this project are presented in Section 3.1.1.

The technical design of the models, including the utility and equipment sizing, will affect the capital and operational cost of each process which will in turn directly affect the profitability of each process. The amount of enzymes used, and ethanol and DDGS produced also differs for different processes and therefore influence the profitability of the plant. Furthermore, the quality of the DDGS produced in the different processes also differs and therefore the price of the produced DDGS differs. Lastly the enzyme dosages and prices also differ for different process models and can also affect the profitability of the process.

3.2.1.2 *Electricity generation*

A CHP plant can be included in the models considered in section 3.1.1 above. This plant will produce electricity and steam for the process. The addition of a CHP plant comes at a cost as it leads to an increase in the CAPEX of the plant. It should be noted that excess electricity is produced in this case which can be sold to the national grid. A process which includes a CHP plant therefore has the advantage of not being dependent on Eskom for power. The modelling of a CHP plant was done in Aspen by using Abdul Petersen (2015) model, and was used to determine the amounts of coal used and electricity produced (Petersen, 2015). In the case of models that used a conventional low pressure boiler (without CHP) for steam production, combined with electricity buy-in from Eskom, the following equation was used to calculate the coal demand for the process.

$$m_{coal} = \frac{m_{water} LHV_{coal}}{\eta_{eff} h_w} \quad [8]$$

Where m is mass, *LHV* is the Lower Heating Value (kJ/kg), η is the boiler efficiency and h_w is the enthalpy of water (kJ/kg).

3.2.1.3 *Coal versus Biomass*

The Overberg and Swellendam region is one of the areas in the WC where there is a potential for triticale-ethanol production, due to proximity to marginal lands suitable for triticale cultivation (Melamu, 2015). Additionally, there are also areas with high concentrations of invasive alien plants (IAPs) in close proximity to Swellendam, which are a suitable fuel source for triticale-ethanol production (Nowell, 2011).

IAPs are more environmentally friendly than straw as IAPs are a threat to local plant ecology and also use higher amounts of water than indigenous plant life (Nowell, 2011). Therefore, IAPs should be cleared. Straw is usually left on fields as it

increases soil fertility and nutrition as well as help retains moisture (Melamu, 2015). This increases the advantage of rather using IAPs, since straw currently has a use while IAPs do not.

In the Overberg region the supply of biomass for process energy is likely to be cheaper than coal. It is cheaper due to the short transport distances in comparison to the transport distance of coal from northern and eastern SA to the WC. Therefore, the price per unit of energy in the form of coal is relatively higher than a unit of energy supplied with biomass, both in the Overberg and other areas of the WC. It should be noted that biomass has a lower energetic value than coal and therefore a larger mass amount of biomass is needed than coal to produce the same amount of steam. Nevertheless, the combined benefits of a lower cost per unit of energy and the carbon neutrality of biomass (which significantly improves the life cycle GHG emission benefits for grain-ethanol production) results in biomass being the preferred fuel for the supply of process energy for grain-ethanol production. The following equation was used to determine the amount of biomass needed for a process being considered:

$$m_{Biomass} = \frac{m_{coal} Cv_{coal}}{Cv_{Biomass}} \quad [9]$$

Where m is mass and *Cv* is Calorific value.

3.2.1.4 *Subsidy*

The subsidy scheme proposed is 15% ROA. The method of depreciation of assets, resulting in the asset values used in calculation of the ROA, is not clearly defined in the Biofuels policy documents from the national government (Department of Energy, 2014a; Department of Minerals and Energy, 2007). The approached followed in this thesis was the straight line depreciation of assets, as is typically used in investment analysis. With the straight line depreciated asset method, the assets are depreciated annual by 6%, until the salvage value is reached in year 17 (2 years construction +15 years plant operation). The equations below are used to calculate the annual subsidy values when the straight line depreciation of assets approach is assumed.

Equation 10, where *n* is the applicable project year, shows how to calculate future inflation in relation to 2015 real values.

$$Inflation_{year\ n} = [Inflation + 1]^n \quad [10]$$

Equation 11 where is used to converts a real cash flow into a nominal cash flow

$$Cash\ Flow\ before\ tax_{nominal} = Cash\ Flow\ before\ tax_{real} * Inflation_{year\ n} \quad [11]$$

Equations 12 and 13 are used to convert asset values into actual nominal straight line

depreciated asset values. Here *i* is the interest/depreciate rate (6%) for the project.

$$Asset\ value_{reduced,real} = Asset\ value_{real} * (1 - i * n) \quad [12]$$

$$Asset\ value_{reduced,nominal} = Asset\ value_{reduced,real} * Inflation_{year\ n} \quad [13]$$

Equation 14 is applied to obtain the straight line depreciated asset value method's ROA.

$$ROA_{straight} = \frac{Cash\ Flow\ before\ tax_{nominal}}{Asset\ value_{reduced,nominal}} \quad [14]$$

Equations 15 and 16 are used together to calculate the straight line depreciated asset value subsidy in real terms.

$$Sub_{straight,\ nominal} = \frac{[15\% - ROA_{straight}] * Asset\ value_{reduced,nominal}}{Ethanol\ L\ produced\ per\ year} \quad [15]$$

$$Sub_{straight,\ real} = \frac{Sub_{straight,\ nominal}}{Inflation_{year\ n}} \quad [16]$$

3.2.2 Economic Assumptions

Below is a list of the economic assumptions made during the process. Further explanations are given after the assumption list below.

- The salvage value is 10% of Fixed Capital Investment (FCI_L).
- The working capital is 10% of FCI_L.
- The plant does not operate for 35 days/year (Amigun et al., 2012).
- The debt to equity ratio is 60:40.
- MACRS depreciation was used for the tax calculation to depreciation assets and was done for a five year period (Amigun et al., 2011).
- The DDGS from the dry grinding process is 61% of the soya oil cake price which is R3935/tonne (Ten Cate, 2015b).
- DDGS from the dry pre-fractionation grinding process is the price of soya oil cake, which is estimated to be R7000/tonne (Ten Cate, 2015a).
- Triticale was assumed to be the same price as B4 wheat (Coetzee, 2015; Ten Cate, 2015b), the minimum price is R1739/tonne, the maximum price is R3212/tonne and hence the average price is R2626/tonne (SAFEX-JSE, 2015).
- CO_2 is not sold.
- The prices of all enzymes, except that of Stargen, are R0.09/ℓ ethanol (Amigun et al., 2011).
- The price of Stargen 002 is R96/kg enzyme (McLean, 2015).

- The price of yeast is R150/kg (Lallemand, 2015).
- The price of coal is R1150/tonne (Metals Consulting International, 2015).
- The price of biomass is R246/tonne (LD Jellison, Inc., 2011).
- The electricity buying price is R0.67/kWh (Eskom, 2015).
- The electricity selling price is R1/kWh (Mashoko et al., 2013).
- The interest rate applied to the bank loan taken out for the plant is assumed to be equal to the prime rate, which was equal to 9.25% nominal for 2014 (South African Reserve Bank, 2015).
- The Target Project IRR for the process is 15% nominal.
- The inflation rate for the project is 5.7% (Statistics South Africa, 2015).
- The 15% ROA calculation is done by using the B4 SAFEX wheat price minus the transport differential (Department of Energy, 2014a).
- Tax is assumed to be equal to 28%, which is tax levied on companies (Amigun et al., 2012; SARS, 2015).
- The plant takes 2 years to construct.
- The plant is operational for 15 years.
- During the first year of operation only 50% of the capacity is achieved, during the second year of production 80% of the capacity is achieved and from the third year of production full capacity of the plant is achieved.
- It is assumed that the boiler capital cost for coal and biomass are the same.

After consultation with the animal feed industry, the DDGS price was assumed to be R3935/tonne for the warm and cold processes (Ten Cate, 2015b). The DDGS from the pre-fractionation dry grinding process could be sold at a price of R7000/tonne; this price was also obtained after consultation with the animal feed industry (Ten Cate, 2015a). The price is motivated by the significantly higher protein and lower fibre content, making it suitable for feeding to non-ruminant animals (e.g. poultry, pigs). Non-ruminant animals feed market is the largest animal feed market in SA (conventional high fibre DDGS can only be consumed by ruminants).

Due to the fact that there is uncertainty regarding the market and price for CO_2, it is assumed that no CO_2 is sold. The buying price of electricity is taken as R0.67/kWh, the value at which it can be brought from Eskom (Eskom, 2015). Economic simulations for a CHP plant were done where surplus electricity is produced and is sold into the grid. The selling price of electricity is taken as R1/kWh (Mashoko et al., 2013). This is a conservative estimate as the electricity that is produced is carbon neutral and therefore can possibly be sold at a higher price.

The triticale feedstock price was assumed to be the same as that of B4 wheat, and the average was taken in real terms from January 2009 to April 2015 resulting in a price of R2626/tonne (Coetzee, 2015; SAFEX-JSE, 2015; Ten Cate, 2015b). This

price is based on communication with KaapAgri and Sjoerd ten Cate. This factory gate price considers the avoided "transport differential penalty" by farmers due to grains being marketed locally, rather than being "exported" north. The local market therefore makes triticale the lowest cost starch grain available in the province. It has a significant cost-advantage over grain sorghum, which is frequently considered as the feedstock for ethanol production elsewhere in SA.

The price of yeast is R150/kg, while the price of conventional "warm" enzymes is assumed to be R115/tonne (R0.09/ℓ) (Amigun et al., 2011). The price of Stargen 002, the enzyme for the cold process, is assumed to be R96/kg (McLean, 2015). Although Stargen 002's price is lower, the fact that it is required in approximately double the dosage as the conventional warm enzymes should be taken into account (Danisco US Inc, 2009).

The cost of coal is assumed to be R1150/tonne, delivered to the factory gate (Metals Consulting International, 2015). Biomass price was taken as R246/tonne, delivered to the factory gate (LD Jellison, Inc., 2011). The price of water for the process is taken as R12/kℓ (Water Rhapsody, 2014).

The debt to equity ratio is 60:40 for the processes' CAPEX. The expected nominal IRR of the project is 15% and the prime bank nominal interest rate is 9.25% (South African Reserve Bank, 2015). As the economic calculation is done in real terms, the inflation value assumed to do future predictions for subsidy calculations is 5.7% (Statistics South Africa, 2015).

3.2.3 Legislation

Legislation stipulates certain conditions in relation to some of the parameters used to model the project. Summarised versions of the legislative requirements in relation to these parameters are relayed in this section.

One of these legislative requirements relates to if ethanol can be sold at the Basic Fuel Price (BFP) for petrol. Ethanol can be sold at the BFP for petrol, provided it is blended at no more than 2% with petrol, which is the mandatory blending for SA (Department of Energy, 2014a). The BFP saw a reduction earlier in 2015 which decreased this value to as low as R4.40/ℓ in February 2015. However, for the base case of this economic analysis in which the assumptions listed in the previous section was used, the average BFP value was used. The average BFP was calculated as the average BFP from January 2009 to April 2015 in real terms and was found to be equal to R6.94/ℓ (Department of Energy, 2014b). The minimum price for the BFP was R3.91/ℓ and the maximum price was R9.21/ℓ (Department of Energy, 2014b).

Another legislative requirement refers to the subsidy calculation. Subsidy needs to be calculated in accordance to a 15% ROA (Department of Energy, 2014a). As seen

in section 2.7.3 the reference grain chosen has an influence on the subsidy calculation. Triticale is the reference grain used for the project in any subsidy calculation. The government, in comparison, uses sorghum as its reference grain for their starch reference plant subsidy calculations (Department of Energy, 2014a).

3.2.4 Sensitivity Analysis

A sensitivity analysis with one variable changing at a time is done for each model to evaluate which model has the highest chance of success. Another reason for doing the analysis is to determine which factors have the largest effect on the profitability of the plant. The factors being varied are:

- Triticale Price;
- Bioethanol Selling Price;
- DDGS Price;
- CO_2 Price;
- Electricity Selling Price;
- Electricity Buying Price;
- CAPEX;
- Coal Price;
- Biomass Price; and
- Number of days when the plant is not operational (capacity).

As only one factor will be varied at a time the other values are kept at their base case values, given in the previous section 3.2.1.4.

3.2.5 Historical Data Analysis

A historical data analysis was done to determine the effect the volatility of the BFP and triticale price have on the profitability of the plant. The ethanol selling price was taken as BFP and the triticale selling price was taken as being equal to the selling price of B4 wheat. All prices were converted into real values applicable on April 2015 by using historical inflation data. The analysis was done from January 2009 to April 2015. Thus, all historical data was converted into actual 2015 values (by inflating them) to evaluate the possible different price variations for BFP and triticale price and their combined effect on the profitability of the plant.

3.3 Environmental Model

In this section the methodology and assumptions for the environmental model will be discussed.

3.3.1 Environmental model methodology

Only one CO_2 balance was done for the best performing economic model, namely model 1 with CHP plant. It was done to compare the environmental effects of using coal versus biomass as energy source for the plant. For this thesis a cradle-to-grave CO_2 emissions approach was considered. In this cradle-to-grave cycle it is assumed that most of the GHG emissions will be CO_2. For this analysis only the nett CO_2 emissions will be considered.

3.3.2 Environmental Assumptions

The environmental assumptions of the CO_2 balance are listed below:

- Agriculture: Triticale production produces 0.2kg of CO_2/kg of grain[20].
- Triticale and biomass transport: It is assumed that a truck runs on diesel and uses 30ℓ/100km[21]. The load capacity of a truck is assumed to be 40 tonne[22]. It is assumed that 1ℓ diesel produces 2.68kg CO_2 [23].
- All the biomass and triticale used for the plant is located within a 100 km radius of the location of the plant[24].
- Coal Transport: The sea freight emission of CO_2 is 124g CO_2/tonne nautical mile[25] (from Richards Bay to Saldanha Bay is 990 nautical miles[26]) and rail freight is 0.026kg CO_2/tonne miles [27] (Saldanha Bay to Swellendam is 332km[28]).
- Production Process: The emission factor is 0.94g CO_2/kg coal[29].
- Ethanol/petrol balance: A car uses 9ℓ petrol/100km[30] and produces 2.35kg CO_2/ℓ petrol[31] and 1.5kg CO_2/ℓ ethanol[32].

In the next chapter the results of the project is presented which was obtained by using the methodology of this chapter.

[20] From (Pineo et al., 2015)
[21] From (Verband der Automobilindustrie, 2012)
[22] From (Verband der Automobilindustrie, 2012)
[23] From (U.S. Energy Information Administration (EIA), 2014)
[24] From (Melamu, 2015)
[25] From (Google maps, 2015)
[26] From(Ports.com, 2014)
[27] From(Immik, 2008)
[28] From (Bureau of Transportation Statistics, 2012)
[29] From (Melamu, 2015)
[30] From (Melamu, 2015)
[31] From (U.S. Energy Information Administration (EIA), 2014)
[32] From (Demirbas, 2008)

4 RESULTS

The methodology set out in the previous chapter was used to model the production of bioethanol using triticale. Models of the four different types of processes, as well as variations of them, that can used to produce bioethanol were built in Aspen and Microsoft Excel. The results obtained from these models are displayed and presented in this section. This was done to give a comprehensive view of the economic feasibility of using triticale for ethanol production in the WC.

4.1 Results from Process Simulations

In this section the process results, namely feedstock, products and utilities of process are presented. In Appendix A the detailed flowsheets as well as stream tables for the individual processes are given.

4.1.1 Feedstock and Products

Table 3 lists the amount of triticale used and the resulting ethanol, DDGS and CO_2 produced from fermentation as per the mass balances calculated from process simulations. At a triticale base feed rate of 407 000 tonne/y for each process, the warm process produces the most ethanol at 160 000 000 l/y, followed by the warm pre-fractionated process, the cold process and the cold pre-fractionated process, at 158 000 000 l/y, 155 000 000 l/y and 154 000 000 l/y, respectively.

TABLE 3: FEEDSTOCK AND PRODUCTS

Models	Warm Process (Model 1)	Cold Process (Model 2)	Warm Pre-Fractionation Process (Model 3)	Cold Pre-Fractionation Process (Model 4)
Triticale (ton/y)	407 000	407 000	407 000	407 000
Ethanol (L/y)	160 000 000	155 200 000	158 400 000	154 100 000
DDGS ton/y	124 600	132 300	66 200	73 100
CO2 (ton/y)	121 900	118 100	120 700	117 300

In terms of the amount of DDGS produced, the cold process produces 132 350 tonne/y, which is more than the warm process (124 600 tonne/y) produces. The warm pre-fractionation process produces the lowest mass of DDGS, 66 200 tonne/y, while the cold pre-fractionated process produces 73 100 tonne/y.

The warm (121 900 tonne/y) and warm pre-fractionation processes (120 700 tonne/y) produces almost the same amount of CO_2 in a year. As was to be expected, the cold process (118 100 tonne/y) produced less CO_2 than the warm process, and the pre-fractionation cold process produced the lowest amount of CO_2, namely 117 300 tonne/y.

4.1.2 Utilities

Table 4 lists the energy, heating and cooling utility demands for the four processes. Each of the demands is based on a triticale feed rate of 407 000 tonne/y for each process, as calculated in the mass & energy balances obtained from the process simulations. The pre-fractionation processes have the lowest heating duty (77.6 MW for warm and 74.9 MW for cold). The warm process, in comparison, has the highest heating duty (86.3 MW) followed by the cold process which has a heating duty of 84.7MW. The specific energy use (MW.y/ℓ ethanol) was lowest for the pre-fractionation processes (0.49 MW.y/ℓ), while the specific energy for the warm process (0.54 MW.y/ℓ) was higher. Finally, the cold process, at 0.55 MW.y/ℓ, was the least efficient at using energy.

For both coal and biomass the annual energy requirements are that the warm pre-fractionation uses the least amount of energy feedstock followed by the cold pre-fractionation then the warm process and with the cold process using the most energy feedstock (See Table 4 for values). The process that has the highest electricity demand is the cold process at 10.2 MW. The electricity demands are as follows: the warm pre-fractionation demand is 9.4 MW, the warm process's is 10.1 MW, and the cold pre-fractionation processes' 9.3 MW. The cooling duty follows the following trend: the warm and cold pre-fractionation is smaller than the cold process, which is smaller than the warm process (73 MW, 79.6 MW and 80.3 MW respectively).

TABLE 4: UTILITIES

Models	Warm Process (Model 1)	Cold Process (Model 2)	Warm Pre-Fractionation Process (Model 3)	Cold Pre-Fractionation Process (Model 4)
Coal (ton/y) with CHP	176 000	173 000	160 000	153 000
Biomass (ton/y) wet	377 000	370 000	343 000	328 000
Coal (ton/y) without CHP	144 000	140 000	130 000	124 000
Biomass (ton/y) without CHP	309 000	301 000	278 000	266 000
Heating Duty (MW)	86.3	84.7	77.6	74.9
Cooling Duty (MW)	80.3	79.6	73.0	73.0
Electricity Needed (MW)	10.1	10.2	9.4	9.3
Electricity Produced (MW)	34.3	33.4	29.8	27.7
Heating Duty (MW)/ethanol (L/y)	0.54	0.55	0.49	0.49

The heating demands of the processes are always met by steam produced onsite in a boiler, whereas the electricity demands can be met either by Eskom-supplied power or by onsite electricity generation. Table 4 indicates the total amount of electricity that can be produced by an onsite CHP plant for the various process scenarios. This electricity will satisfy the demand for electricity for the process while in addition providing surplus electricity for export to the national grid.

The sizing of the CHP plant is done according to the process steam demand, which is associated with a particular amount of electricity production. The latter is often in excess of the process energy demands. The amount of (surplus) electricity produced

by CHP plants followed the same trend as the heating duty, since steam and electricity are always co-generated in a CHP plant: A higher heating demand will require more steam and therefore automatically produce more (surplus) electricity. It should be noted that a CHP plant uses more coal/biomass (19%) than if the coal/biomass was used for PH. This is true, since some of the energy in the CHP plant is used for the production of electricity. The amount of biomass needed for the PH is 53% more than the amount of coal required to achieve equivalent energy inputs to the boilers for PH.

4.2 Economic Results

In this section the CAPEX, OPEX, base case subsidy and IRR obtained for the different processes is presented. The results of the historical data analysis and sensitivity analysis are also shown in this section. In Appendix B, additional graphs (not included in this section) illustrating the effect of different subsidies are presented. Additional IRR graphs are also shown in Appendix C. The subsidy being considered was calculated using the reference grain as triticale rather than sorghum for the 15% ROA calculation.

4.2.1 CAPEX and OPEX

The capital (CAPEX) and operating (OPEX) costs of the four processes, including the options with a CHP plant for onsite electricity generation, are given in Table 5. The CAPEX values for options that include CHP plants, in increasing order, are as follows: the warm process (R3 221 M), cold process (R3 225 M), warm pre-fractionation (R3 699 M) and cold pre-fractionation (R3 904 M). The CAPEX values without the CHP plant included follow a similar trend. Although these CAPEX values are significantly lower than the options that include a CHP plant, the difference can be ascribed to the difference in the amount of CAPEX between a low pressure PH and a CHP plant. In increasing order the CAPEX value for the CHP plant options is as follows: the warm process (R2 274 M), cold process (R2 356 M), warm pre-fractionation (R2 765 M) and cold pre-fractionation (R2 784 M). Therefore, the CAPEX increases by 25-29% when an onsite CHP plant is included in comparison to the option when only a low pressure PH is installed.

The above CAPEX values can be evaluated by the CAPEX requirement per litre ethanol produced annually. The CAPEX/ℓ for the CHP plant processes is R25.3/ℓ for the cold pre-fractionation, R23.4/ℓ for the warm pre-fractionation, R20.8/ℓ for the cold process and R20.1/ℓ for the warm process. CAPEX/ℓ for the without CHP plant processes is lower than for the with CHP plant processes due to lower equipment costs. It is R18.1/ℓ for the cold pre-fractionation, R17.5/ℓ for the warm pre-fractionation, R15.2/ℓ for the cold process and R14.2/ℓ for the warm processes.

Brain Tait's value for a sorghum ethanol plant was R15.19/ℓ in 2012 (Tait, 2014, 2011). When this value is scale up to 2014, this price increases to R23/ℓ which is higher than the calculated CAPEX/ ℓ for the plants with PH in this project.

TABLE 5: CAPEX AND OPEX

Models	Warm Process (Model 1)	Cold Process (Model 2)	Warm Pre-Fractionation Process (Model 3)	Cold Pre-Fractionation Process (Model 4)
CAPEX (R Million)	R 3 221	R 3 225	R 3 699	R 3 904
CAPEX (R Million) without CHP	R 2 274	R 2 356	R 2 765	R 2 784
OPEX (R Million)	R 1 549	R 1 589	R 1 533	R 1 564
OPEX (R Million) without CHP	R 1 566	R 1 606	R 1 543	R 1 580
CAPEX/L ratio	20.1	20.8	23.4	25.3
CAPEX/L ratio without CHP	14.2	15.2	17.5	18.1
OPEX with Biomass (R Million)	R 1 440	R 1 482	R 1 433	R 1 469
OPEX with Biomass (R Million) without CHP	R 1 476	R 1 519	R 1 467	R 1 503
OPEX/L ratio	9.68	10.24	9.68	10.15
OPEX/L ratio without CHP	9.79	10.35	9.74	10.25
OPEX/L ratio Biomass	9.00	9.55	9.05	9.53
OPEX/L ratio without CHP Biomass	9.23	9.78	9.26	9.75

The annual OPEX values in increasing order for the process options with CHP plants were as follows: the warm pre-fractionation process (R1 533 M), warm process (R1 549 M), cold pre-fractionation process (R1 564 M) and cold process (R1 589 M). This successive trend observed for the OPEX of processes with CHP plants was also followed for the processes without CHP plants. The only difference in the observed trend for the processes with CHP plants was that their OPEX values were approximately 1% larger than those of the processes without CHP plants. The annual OPEX values in increasing order for process options without CHP plants were as follows: Warm pre-fractionation process (R1 543 M), warm process (R1 566 M), cold pre-fractionation process (R1 1580 M) and cold process (R1 606 M).

The OPEX for plants with CHP, on the basis of Rand per litre ethanol were: R9.68/ℓ for the warm and warm pre-fractionation, R10.25/ℓ for the cold pre-fractionation and R10.35/ℓ for the cold process. These R/ℓ values were slightly higher for plants without CHP. In many cases the OPEX values were therefore higher than the ethanol selling prices (e.g. base case BFP of R6.94/ℓ), indicating that additional product sales are required to render the process profitable.

The OPEX when biomass instead of coal was used as fuel to generate energy for the process was 4.9-7.1% lower. Besides this decrease in OPEX when biomass is used, it followed a similar trend as the trend observed for the OPEX when coal was used. The difference between the OPEX with CHP plants and without CHP plants is slightly wider for biomass (2.5%) than for coal (1%). Thus, the lowest value in R/ℓ for ethanol price, when only considering ethanol sales, was R9/ℓ for the warm process with a CHP plant using biomass as energy source.

The OPEX can be broken downed into the cost of raw materials (CRM), the cost of utilities (CUT), the cost of operating labour (COL) and the maintenance repairs and insurances. Figure 14 shows the breakdown of the OPEX for model 1 with CHP. The reason for choosing this model as an illustration of the OPEX breakdown will be discussed later in section 4.2.2. 75% of the OPEX are from raw materials, which in

this case is the cost associated with feedstock procurement Utilities also comprises a major part of OPEX but to a lesser extent than the feedstock (15.8% versus 75%).

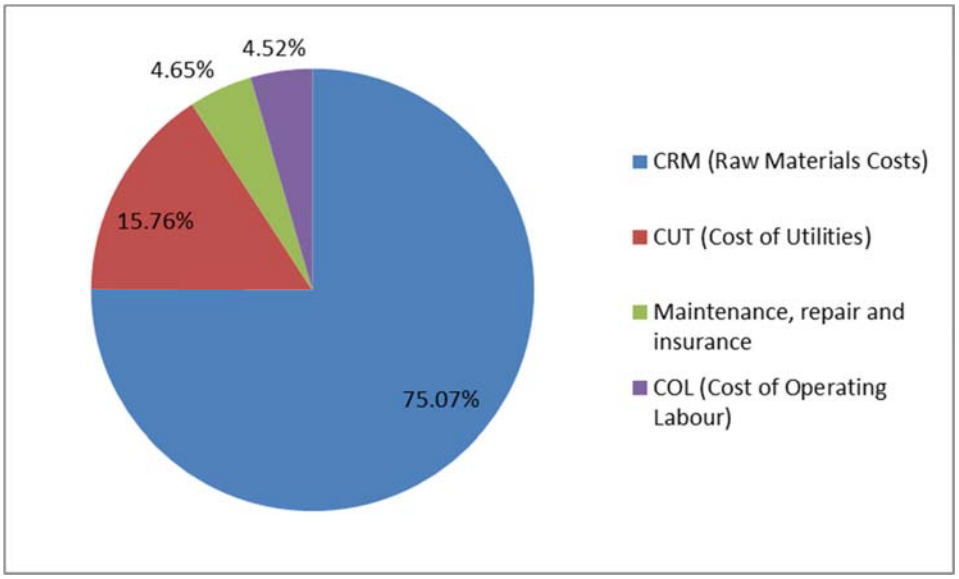

FIGURE 14: BREAKDOWN OF OPEX FOR MODEL 1 WITH CHP

4.2.2 Subsidy and IRR

Throughout all of the options the IRR (see Table 6) for the warm process is significantly higher than for the other processes. The slightly lower IRR for the "CAPEX included CHP plant with coal as an energy source", compared to the use of biomass (CAPEX included CHP plant) is due to a lower biomass cost (Table 6). Therefore, the model with the highest IRR (4.79 %) was the warm process with a CHP plant that uses biomass as its fuel source.

TABLE 6: BASE CASE IRRS

Models	Warm Process (Model 1)	Cold Process (Model 2)	Warm Pre-Fractionation Process (Model 3)	Cold Pre-Fractionation Process (Model 4)
IRR (CAPEX included CHP plant, coal as energy source)	3.02%	2.40%	2.15%	1.90%
IRR(CAPEX included CHP plant, Biomass as energy source)	4.79%	3.85%	3.17%	2.66%
IRR (CHP plant not included in CAPEX, using coal as energy source)	1.56%	1.56%	1.56%	1.56%
IRR (CAPEX excluded CHP plant, biomass as energy source)	2.31%	1.76%	1.79%	1.68%

To cover the debt payments of the process, a real project IRR of 3.5% needs to be achieved. The expected real project IRR of the process is 9.7%. An IRR of at least this level should be achieved to ensure that the investors receive their desired return on investments. It is also of interest to note that for the base case of economic assumptions, only two processes achieved an IRR that covered the debt payment (warm and cold with CHP using biomass as energy source) and none of the

processes reached the investor's expected project IRR of 9.3% (real terms). The IRRs for processes excluding a CHP plant (all at 1.56%), were lower than the IRRs for the same processes that included a CHP plant, e.g. IRRs of 3.02%, 2.40%, 2.15% and 1.90% were achieved.

Table 7 shows the required subsidies for all the processes and different scenarios. Note that the "subsidy" was calculated as the rand per litre ethanol (R/ℓ) value required from the government to achieve a 15% ROA using triticale as reference grain. It should be noted that the values given in Table 7 are for the first year the plant was fully operational.

TABLE 7: BASE CASE SUBSIDIES

Models	Warm Process (Model 1)	Cold Process (Model 2)	Warm Pre-Fractionation Process (Model 3)	Cold Pre-Fractionation Process (Model 4)
Subsidy (CAPEX included CHP plant, coal as energy source)	R 1.00/L	R 1.36/L	R 1.71/L	R 2.10/L
Subsidy(CAPEX included CHP plant, Biomass as energy source)	R 0.32/L	R 0.67/L	R 1.08/L	R 1.48/L
Subsidy (CHP plant not included in CAPEX, using coal as energy source)	R 1.91/L	R 1.92/L	R 2.02/L	R 2.21/L
Subsidy (CAPEX excluded CHP plant, biomass as energy source)	R 0.97/L	R 1.36/L	R 1.54/L	R 1.71/L

There is no process that does not require a subsidy in order to achieve a 15% ROA if only ethanol and DDGS sales are considered as revenue. The lowest subsidy requirement according to the 15% ROA methodology was R0.32/ℓ for the warm process with a CHP plant that used biomass as its choice of fuel source. All of the processes followed similar trends as those observed in Table 6, and the warm process consistently received lower subsidies than the other processes. It should also be noted that the use of biomass as fuel source and the inclusion of a CHP plant also lowers the subsidy requirement.

Figure 15 shows how the required amount of subsidy decreases yearly for model 1 with a CHP plant, which was the model that performed best. During the first two years no subsidy needs to be given as the plant is being built and therefore no sales are made. During the next two years the size of subsidy needs to be large since the plant is not yet fully operational. Accordingly, during year 2 (50% of capacity) the subsidy is R4.20/ℓ and during year 3 (80% of capacity) it is R1.85/ℓ. The subsidy systematically decreases until R0/ℓ is reached at the end of year ten. It should be noted that even though zero subsidy is achieved after ten years, this result is based purely on the comparison of annual EBIDTA (Earnings before Interest, Tax, Depreciation and Amortisation) values to asset values (after depreciation). As a result the models still do not achieve acceptable IRRs to cover the debt payments and/or meet investors' expectations (3.05% versus 3.5%).

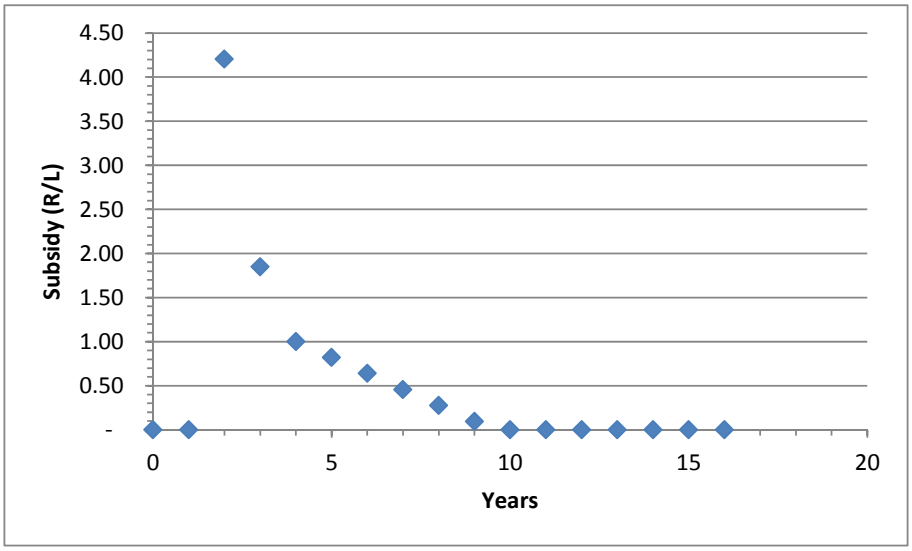

FIGURE 15: SUBSIDY VERSUS YEARS FOR MODEL 1 WITH CHP

4.2.3 Historical Prices Analysis

In this section, historical data is used in the various models that were built, to simulate the real-time (2015) effect of price variation on the considered plant's profitability. The methodology used to obtain these 2015 values using historically adjusted values is discussed in section 3.2.5. In Figure 16 the historical corrected price over the term January 2009 to April 2015 are illustrated for the various feedstock prices and BFP that are considered.

Both the B4 wheat and sorghum price experience have been rather volatile over the observed 6 year period. The B4 wheat price is taken as the price of triticale since triticale is not currently sold commercially. It is also of interest to note that the B4 wheat and sorghum prices often move in opposite directions. This could be ascribed to the fact that these two types of grain do not grow in the same area and therefore different external factors can influence them.

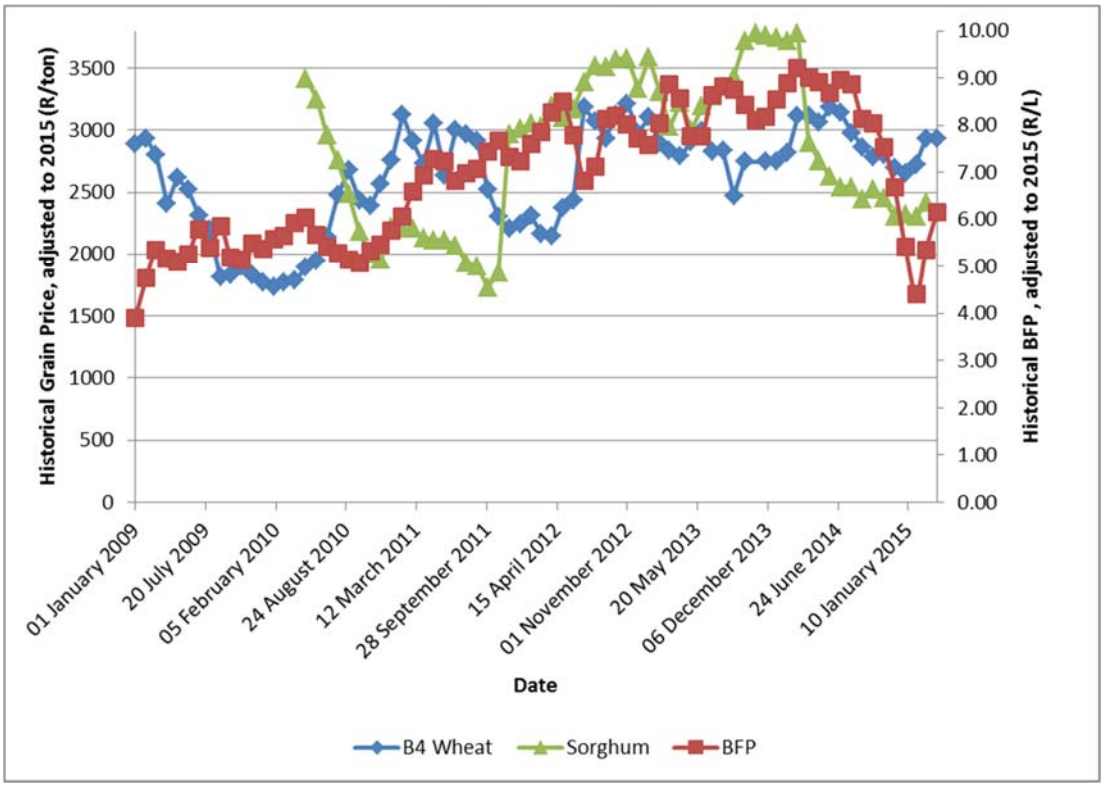

FIGURE 16: HISTORICAL CORRECTED PRICES FOR GRAIN AND BFP

All of the considered grain prices show an upwards trend price. The graph also shows a significant decrease in the BFP during 2013. This decrease could stem from the fact that the BFP is influenced by the dollar rand exchange rate as well as by the price of crude oil.

Figure 17 shows the relationship between BFP and triticale prices. According to the graph there is a very weak positive correlation as the R^2 value is small (0.18) and the data seems randomly scattered. The weak relationship indicates that there is only a small interdependency between the BFP and price of grain and therefore an increase in the BFP price might result in an increase in the grain price. Accordingly, to simulate the combined effect variations in the BFP and triticale price have on the profitability of the different models, historical data was adjusted to 2015 prices.

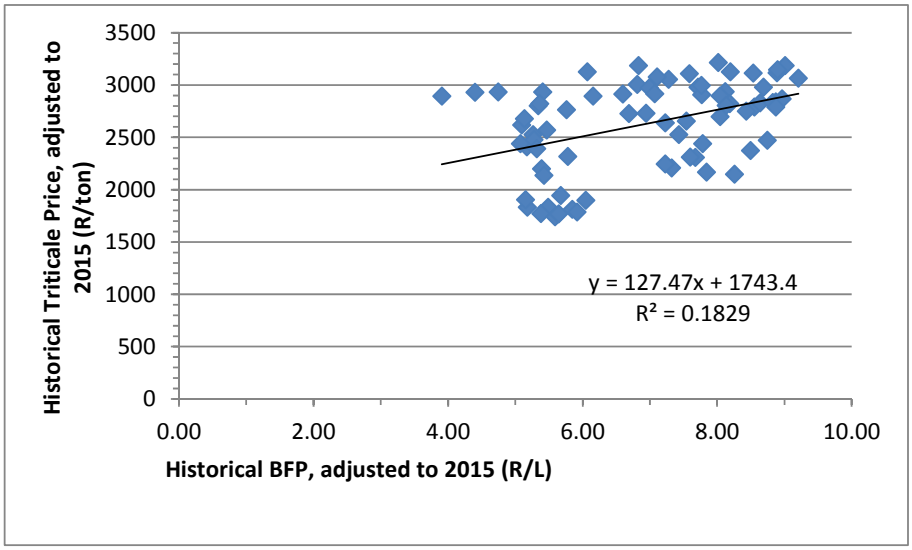

FIGURE 17: RELATIONSHIP BETWEEN BFP AND TRITICALE PRICES

From Figure 18 it is clear that the IRR experienced large variations in its value as a result of historical price variations in the BFP and assumed triticale prices (see Figure 16). The model that performs the best as well as achieves the highest IRR (11%) is the warm process (model 1) with CHP plant. The process that performs the worst is the cold pre-fractionated process (model 4) without CHP plant, with a maximum IRR of 7.34%. Therefore, the selling of electricity (CHP plant) is the predominant factor influencing IRR at low BFP and high triticale prices (unfavourable conditions). Products and feedstock prices, in comparison, are the predominant factors that influence the IRR of the process at high BFP and low triticale prices (favourable conditions).

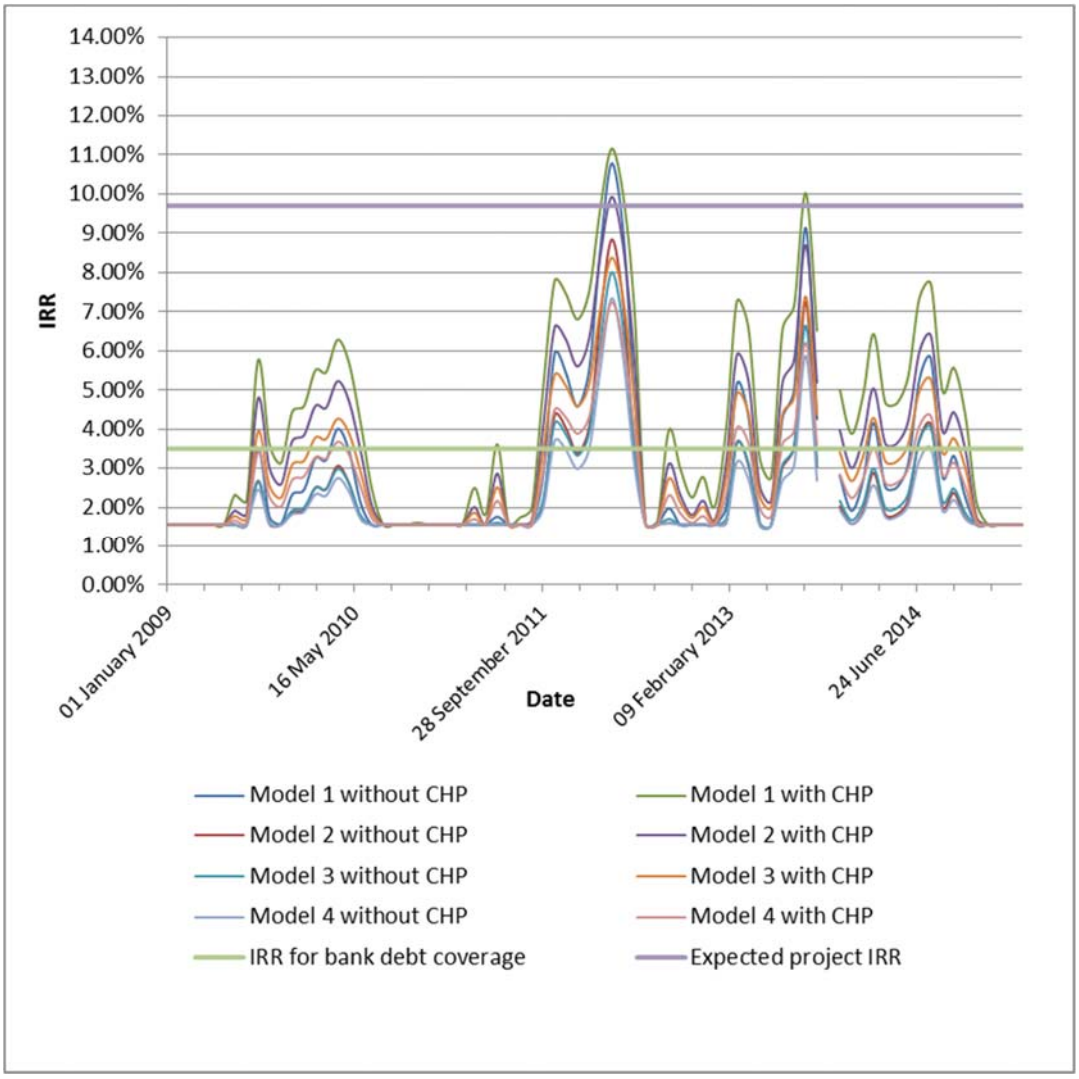

FIGURE 18: HISTORICAL DATA USED FOR IRR CALCULATION USING COAL

Figure 19 shows the frequency with which the project IRR in Figure 18 is below the debt payment IRR (3.5%). It also shows the frequency with which the project IRR is

above the debt payment IRR (3.5%) but still below expected project IRR (9.7%), and finally also the frequency with which it is above the expected IRR (9.7%). All of these calculations were made in relation to model 1 (warm process) with a CHP plant using coal. As can be seen from Figure 19 the IRR was 48% of the time too low to pay back the project's debt. 48% of the time the IRR was high enough to indicate that the debt repayment could be made. Only 4% of the time the expected project IRR desired by investors would be achieved and therefore this may seem like a risky investment for investors.

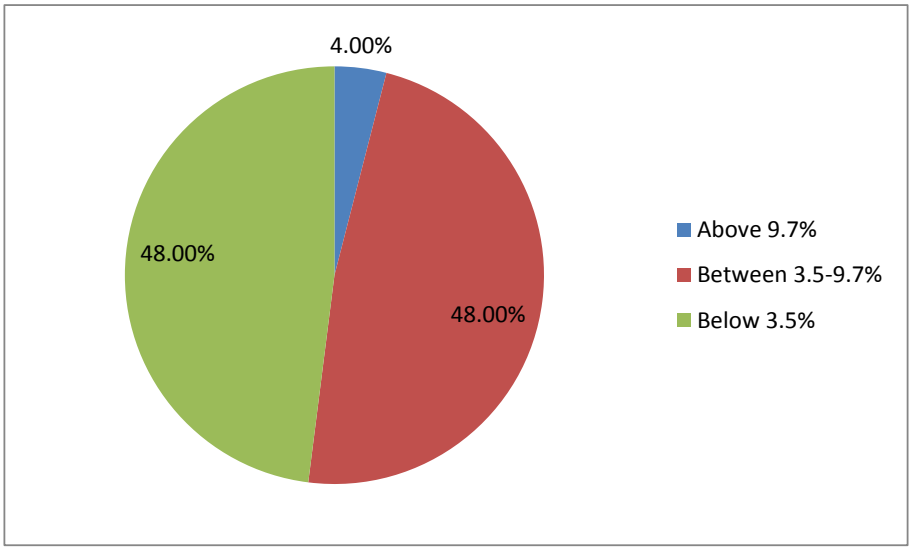

FIGURE 19: IRR FREQUENCY HISTORICAL RANGES FOR MODEL 1 WITH CHP USING COAL

Figure 20 indicates the same results as those acquired in Figure 18 except that the IRRs are higher in Figure 20. The IRRs are higher due to the process energy demand being met by using biomass rather than coal. Biomass is cheaper in the WC than coal. It is evident, due the observed higher IRR values, that the expected project IRR is more likely to be achieved. Furthermore, the debt payment is also more likely to be made due to the higher IRR values. More detail about the difference in the process' project IRR when using biomass in comparison to coal is shown in Figure 21 below for model 1 with a CHP plant.

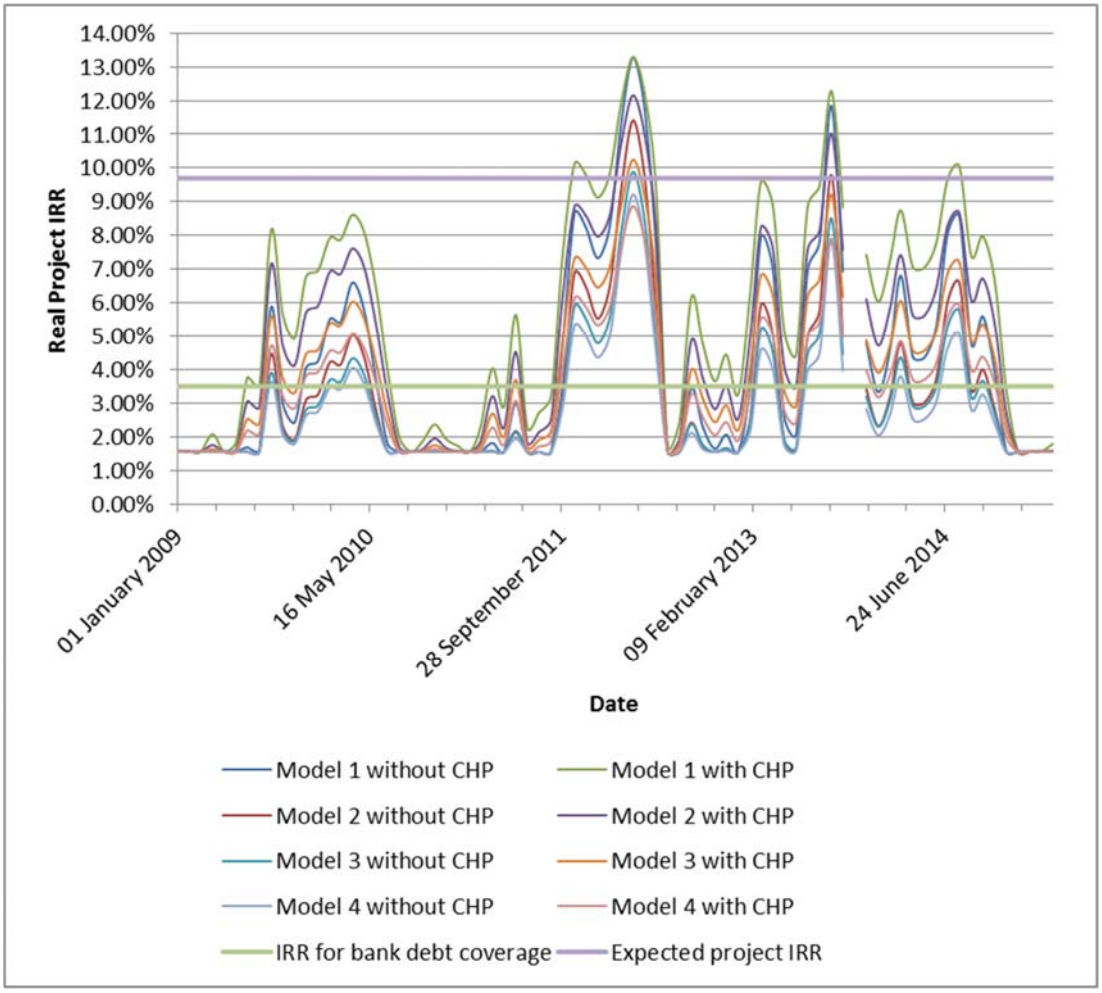

FIGURE 20: HISTORICAL DATA USED FOR IRR CALCULATION USING BIOMASS

Figure 21 shows the effect that historical prices have on the IRR of model 1 (warm process) with a CHP plant using coal, and on model 1 with a CHP plant using biomass. The BFP and triticale prices vary greatly and thus the IRR values are varied. The process using biomass constantly outperforms the process using coal. The highest IRR for biomass is 13.7% while the highest project IRR for coal is 11.6%. Most of the time a large enough IRR was achieved to cover the debt payment, while only in very favourable conditions a large enough IRR was achieved to meet the investors' expected project IRR.

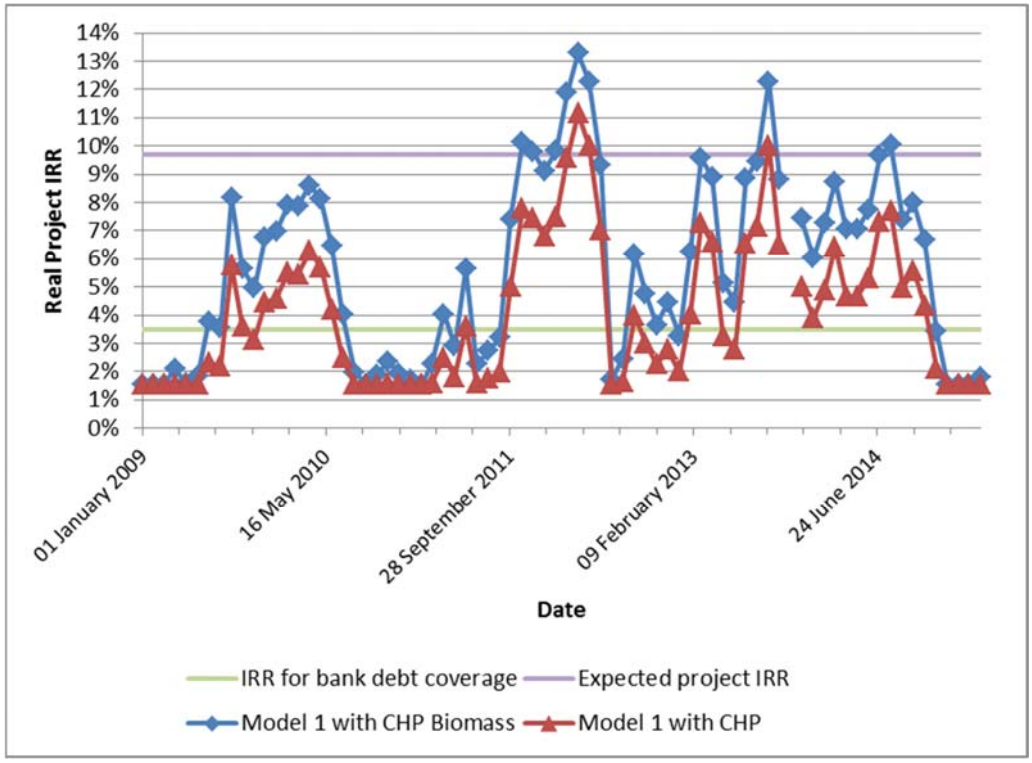

FIGURE 21: BIOMASS AND COAL IRR FROM HISTORICAL DATA FOR MODEL 1 WITH CHP

Figure 22 shows the relationship between the IRR and subsidy obtained for all of the models. As depicted in the figure a definite relationship can be seen between these two variables. The IRR is dependent on the subsidy in the range R0.01-R2/ℓ. At subsidies close to zero the IRR is independent of subsidy. The IRR reaches a minimum value at subsides larger than R2/ℓ and stays constant thereafter. At these constant levels the subsidy is independent of the IRR. It was further observed that when the subsidy was high, the IRR was low and when subsidy was low, the IRR was high. This is desirable as it encourages the most profitable cases of grain to ethanol plants without being dependent on a high subsidy.

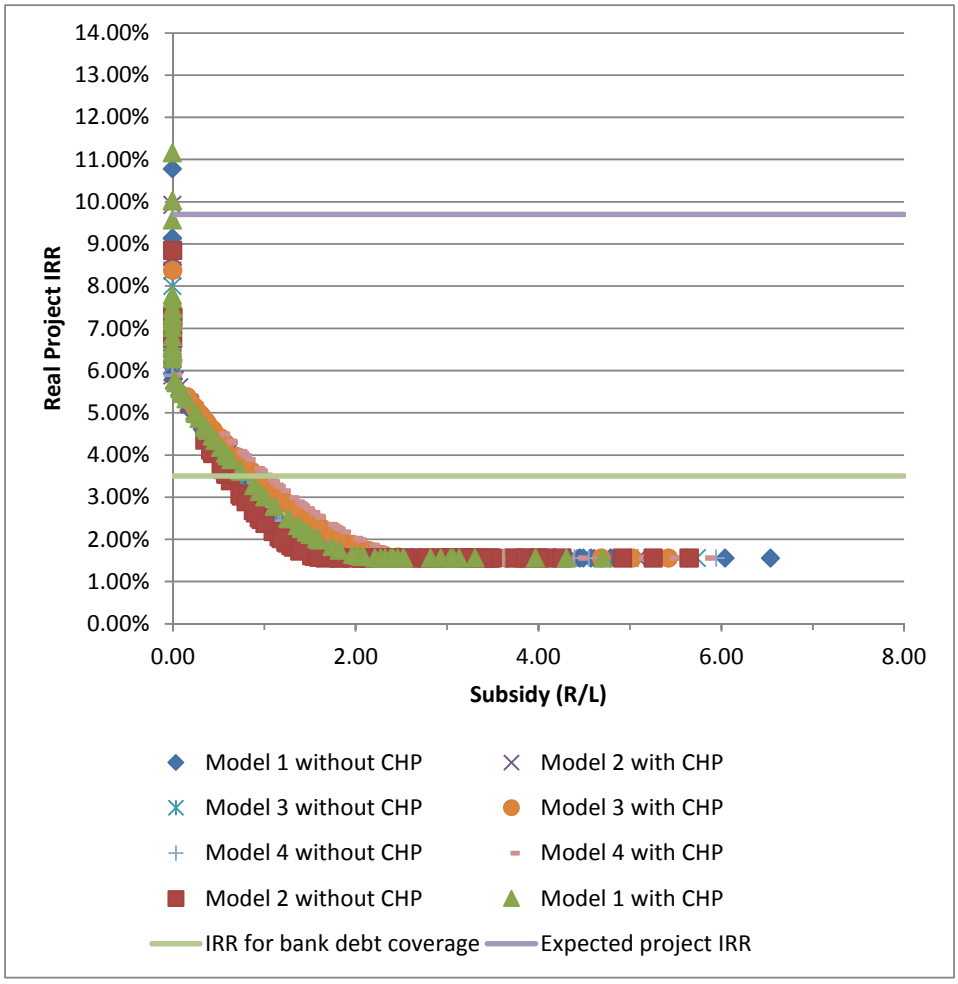

FIGURE 22: IRR VERSUS SUBSIDY

In the middle range there is a negative correlation between the IRR and subsidy. Also of interest is the fact that the minimum IRR does not cover debt payment regardless of the amount of subsidy given. Furthermore, it is observed that the expected project IRR is not always achieved for a subsidy of zero.

Figure 23 shows the effect of using the SAFEX sorghum prices in the subsidy

calculation that is calculated in terms of the 15% ROA methodology, versus using the assumed triticale grain prices. The model that is used for Figure 23 is model 1 (warm process) with a CHP plant as it has the highest IRR of all the models. Figure 23 shows clearly that by using sorghum as references grain, higher IRRs can be achieved when sorghum is sold at a higher price than triticale. The expected project IRR and the IRR required for the debt coverage are more often achieved when using sorghum as the references grain than when using triticale as the reference grain of the plant. It is also clearly shown that when the sorghum price is lower than the triticale price, negative IRRs can be achieved. When using triticale as references grain, negative IRRs are not obtained as the models have a minimum IRR of 1.56%. It should be noted that using sorghum as reference grain in the base case leads to achieving a larger IRR (5.54% versus 3.05%) than for when triticale is used.

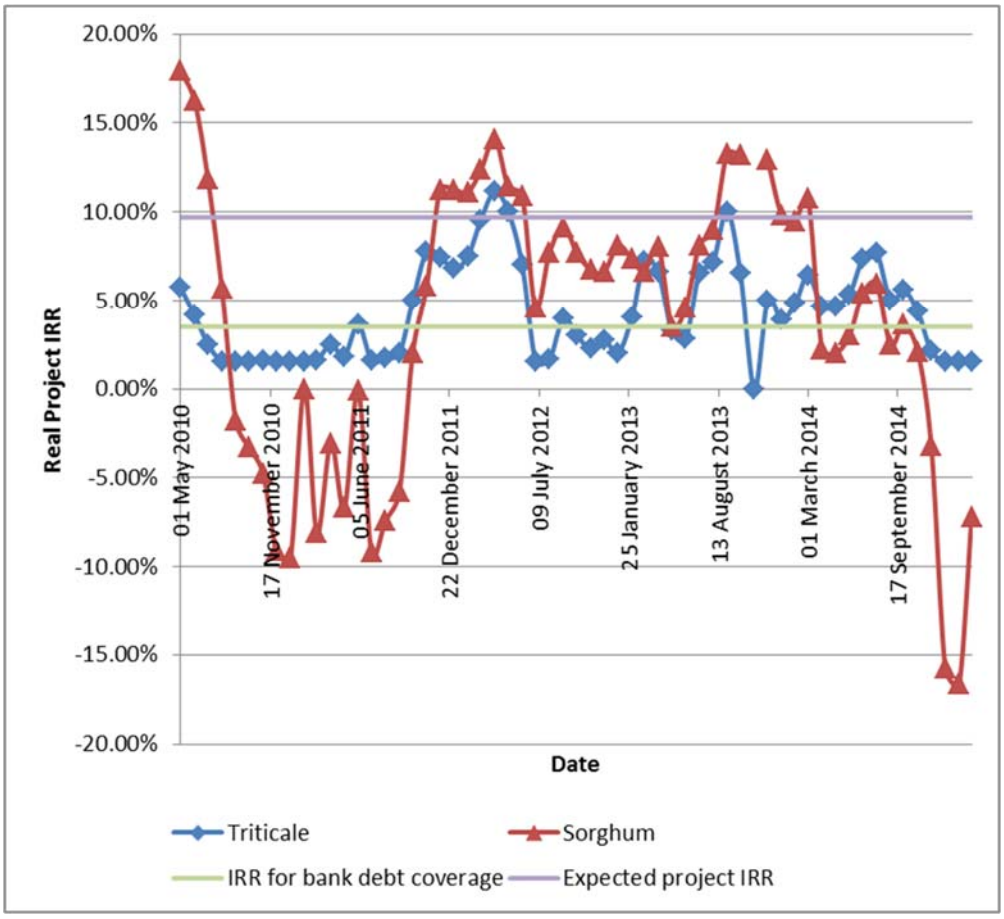

FIGURE 23: EFFECT OF USING SORGHUM AS REFERENCES GRAIN TO CALCULATE SUBSIDY

Figure 24 shows the relationship between triticale and the BFP for different IRR ranges. The model used for this graph is the model with the highest IRR, namely model 1 (warm process) with a CHP plant using biomass as its energy source. It is clear from Figure 24 that only in 5.33% of the cases the expected real IRR (9.7%)

was achieved. Furthermore, the debt payment IRR was only achieved in approximately half of the cases (53.33%). In 41.33% of the cases the IRR was not high enough to reach the debt payment IRR of 3.5%.

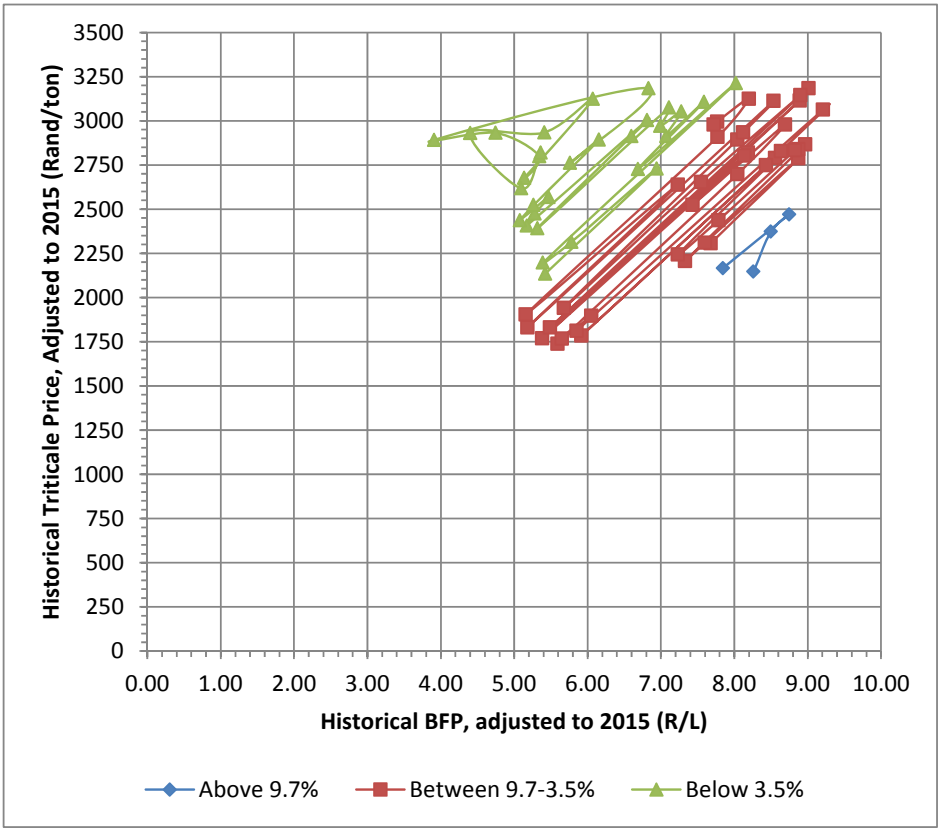

FIGURE 24: RELATIONSHIP BETWEEN TRITICALE AND BFP FOR DIFFERENT IRR GROUPS

There are definitely distinct bands of BFP and triticale prices associated with certain ranges of IRR. If the BFP value is low (R4-5/ℓ) but the triticale price is above R2000/ton the project IRR is too low to cover the debt payment. Even for high BFP prices the debt payment cannot be made if the triticale price is too high (±R3200/ton). The investors' expected project IRR was only achieved at high BFP (above R7.5/ℓ) and low triticale prices (below R2000/ton). Hence, the range between the expected project IRR and the IRR required for debt coverage will be achieved in the mid range values of the BFP and triticale price. Therefore, to the render the project profitable, the BFP should be high and the triticale price should be low. If the BFP and triticale price were known, it is possible to determine in which IRR band the values fall and if the plant will be profitable at the specified prices.

4.2.4 Sensitivity Analysis

A sensitivity Analysis was done for the different factors influencing the profitability of the plant. Presented in this section are the results of the sensitivity analysis for these factors namely the BFP, triticale price, DDGS price, CO_2 price, coal price and biomass price, electricity price, capacity, and CAPEX. These values were varied relative to their base case assumptions, to identify input values that would result in acceptable project IRRs, to meet investor expectations. The range for the BFP and triticale price was determined from the previous section's historically adjusted prices that were adjusted to 2015.

Figure 25 summarises the effects of different factors on the IRRs achieved when using the warm process (model 1) with a CHP plant. The figure is a summary of part of the findings obtained (based on ±20% variation in parameters) from the sensitivity analysis results below. The factor that had the largest influence on the IRR of the plant was the BFP (5.69%) followed closely by triticale price (5.50%). Ethanol producers may have some control over the triticale price through having negotiations with farmers supplying grain to production facilities. In comparison, the exposure to the BFP cannot be mitigated in any way, as it is dependent on a range of uncontrollable local and international factors (such as crude oil price).

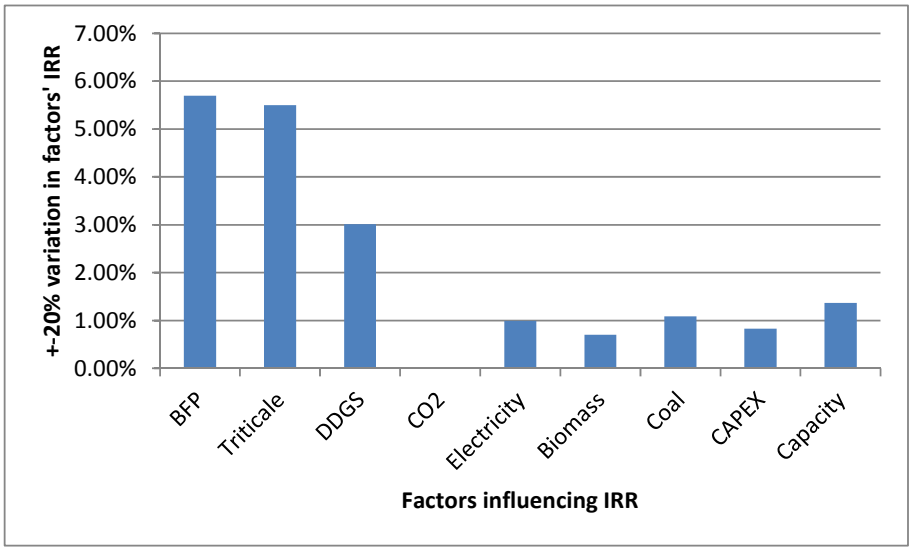

FIGURE 25: EFFECT OF IRR ON DIFFERENT FACTORS 20% VARIATION

DDGS also has an effect on the project IRR but to a lesser extent. This can be seen, as the 20% variation analysis on the DDGS causes only a 3.01% variation in the IRR relative to an approximate 6% variation caused when variating the BFP or triticale price. This 3% variation (caused by variating the DDGS price) would still have an influence on the profitability of the plant and therefore an accurate DDGS price is important. The other factors considered in the sensitivity analysis have a minimal

influence (±1% variation in the project IRR) on the profitability of the plant.

4.2.4.1 *BFP*

The effect of BFP on the project IRR is illustrated in Figure 26. It indicates a minimum IRR value of 1.56% that present itself as a plateau of values under economically unfavourable circumstances (very low BFPs). As the BFP increases, the IRR of the process also increases (positively correlated). It is also illustrated in Figure 26 that the highest project IRR is achieved when using the warm process with a CHP plant (Model 1 with CHP). The resulting IRR value ranges from 1.56% at a BFP of R3.91/ℓ (minimum price) to 10.25% at a BFP of R9.21/ℓ (maximum price).

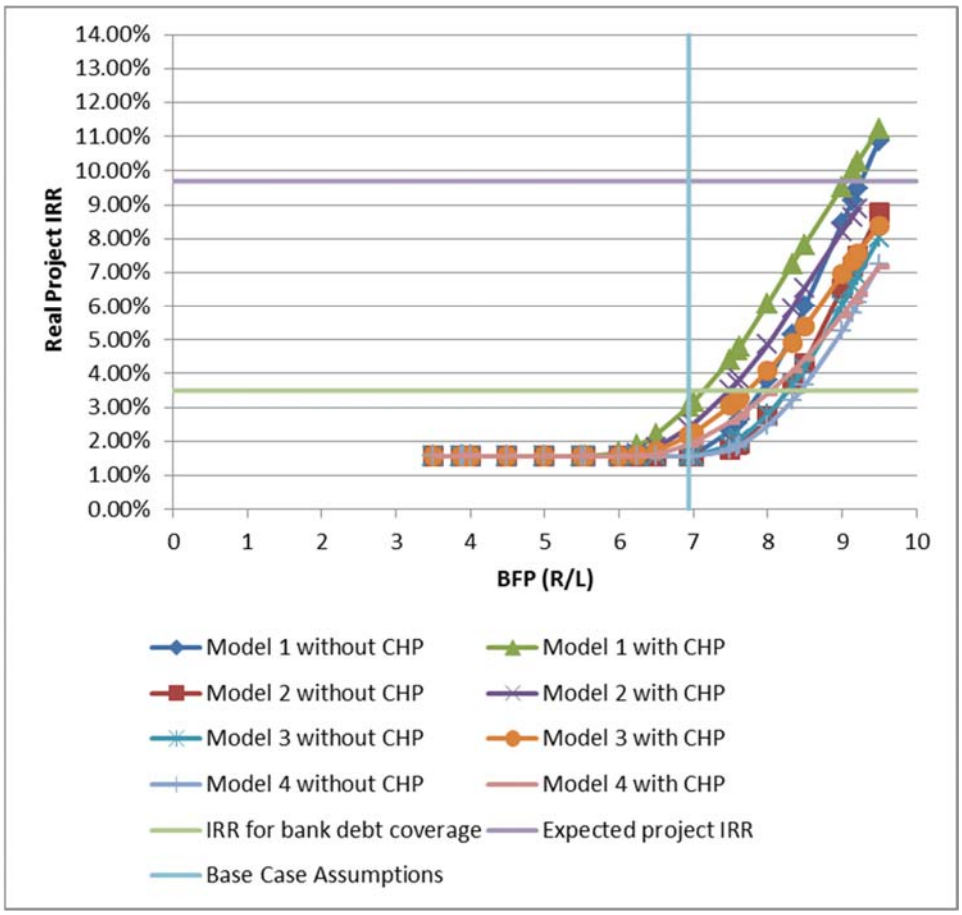

FIGURE 26: BFP VERSUS IRR

For all the processes, including the warm process, the sale of surplus electricity produced onsite by the CHP plant improved the profitability of the process. This result was obtained in comparison to scenarios without a CHP plant where electricity has to be brought-in from Eskom.

The IRRs increased from 1.56% (model 1 without CHP plant) to 3.02% (model 1 with

CHP plant) at a BFP of R6.94/ℓ (base case value) and continued this upwards trend as the BFP was increased. The warm process performed better than the cold process (model 2) in terms of its IRR's performance, which indicates that the warm process required less subsidy.

Consideration should be given to experience relating to the threshold IRR values of 3.5% and 9.7%. Below an IRR of 3.5 % the bank debt cannot be repaid, while below 9.7% investors will not receive their expected returns on investment. The project IRR (9.7% in real terms) is achieved in very few cases (only a BFP above R9.05/ℓ for the warm process with a CHP plant and a BFP above R9.25/ℓ for warm process without a CHP plant). If the cold pre-fractionated process simulated in model 4 (the worst performing model), was to achieve the expected project IRR, the BFP must be at least R10.15/ℓ. This high BFP prices is not ever reached in the historical BFP price range and therefore is unlikely to be attained. To achieve the IRR (3.5%) needed to cover the debt payment, the minimum required BFPs were R7.15 to R8.44/ℓ from best to worst performing model. It is also of interest to note that at a low BFP, the cold process with a CHP plant achieves a higher IRR than the warm process without a CHP plant. This connection between the BFP and the IRR of the different models changes at the upper end of the BFP range.

In Figure 27 the effect of the different triticale and DDGS base cases prices (initial values) have on the IRR is shown. The effect on the project IRR of the two models that performed best, namely Model 1 with a CHP plant and model 3 with a CHP plant, was shown. These models performed the best for each set of initial values. The SAFEX values used in this analysis were R2626/tonne for triticale, R3935/tonne for non-pre-fractionated DDGS and R7000/tonne for pre-fractionated DDGS. These values were chosen as the base case values for the whole project. The alternative values that were also considered were R2000/tonne for triticale, R1800/tonne for non-pre-fractionated DDGS and R7000/tonne for pre-fractionated DDGS. The graph clearly shows that model 1 (the warm process) with a CHP plant is not as greatly influenced as might be expected, since the lower triticale price is mitigated by the higher DDGS price.

This is not true for model 3 (the warm pre-fractionation process) with a CHP plant, since it surpasses the other models due to its lower triticale price. Therefore it is observed to achieve the expected project IRR (R7.80/ℓ) and the IRR needed for debt coverage (R5.90/ℓ) at a lower BFP than the SAFEX method. A low triticale price therefore has a dominant effect on the economic viability, compared to the influence of variations in the DDGS prices.

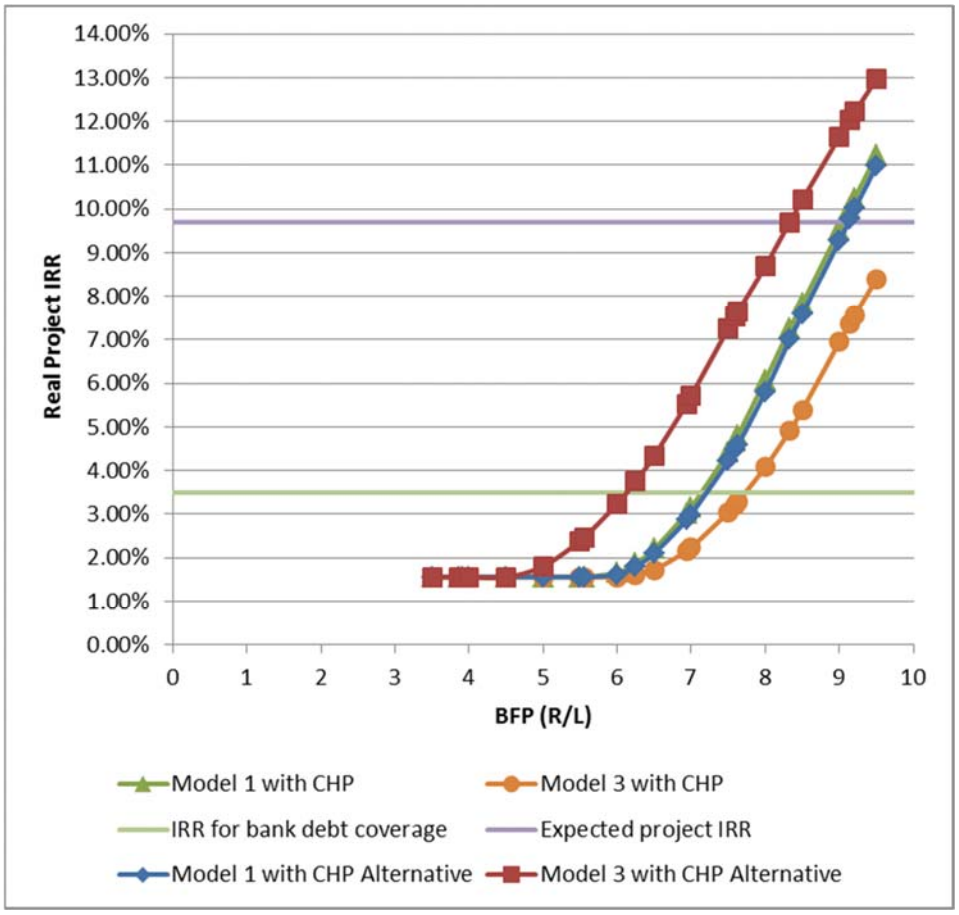

FIGURE 27: THE EFFECT OF DIFFERENT INITIAL PRICE VALUES ON IRR

Currently when using the values for the base case models, the DDGS price obtained in the pre-fractionated process versus the price obtain when the process used was not pre-fractionated, is only 1.8 times bigger. This price difference is not big enough to render the additional CAPEX needed for the pre-fractionated processes profitable, as can be seen in Figure 27 where the base case model 1 outperformed model 3. When using the alternative values method for calculating the DDGS price, model 3 outperforms model 1 due the pre-fractionated DDGS price being 3.9 times bigger. Thus, to render the pre-fractionated processes more profitable than the non pre-fractionated processes, the DDGS price should be at least 2.5 times higher for the pre-fractionated DDGS than the non pre-fractionated DDGS price.

4.2.4.2 *Triticale Price*

Figure 28 shows an overall downward trend in the IRR and economic viability as the price of triticale increases, until it stabilises at an IRR of 1.56%. The most profitable process (as illustrated in Figure 28) is the warm process (model 1) with a CHP plant, followed by the warm process without a CHP plant, then the cold (with a CHP plant),

cold (without a CHP plant), warm pre-fractionation (with a CHP plant), warm pre-fractionation (without a CHP plant), cold pre fractionated (with a CHP plant), and finally, the cold pre fractionated (without a CHP plant) processes.

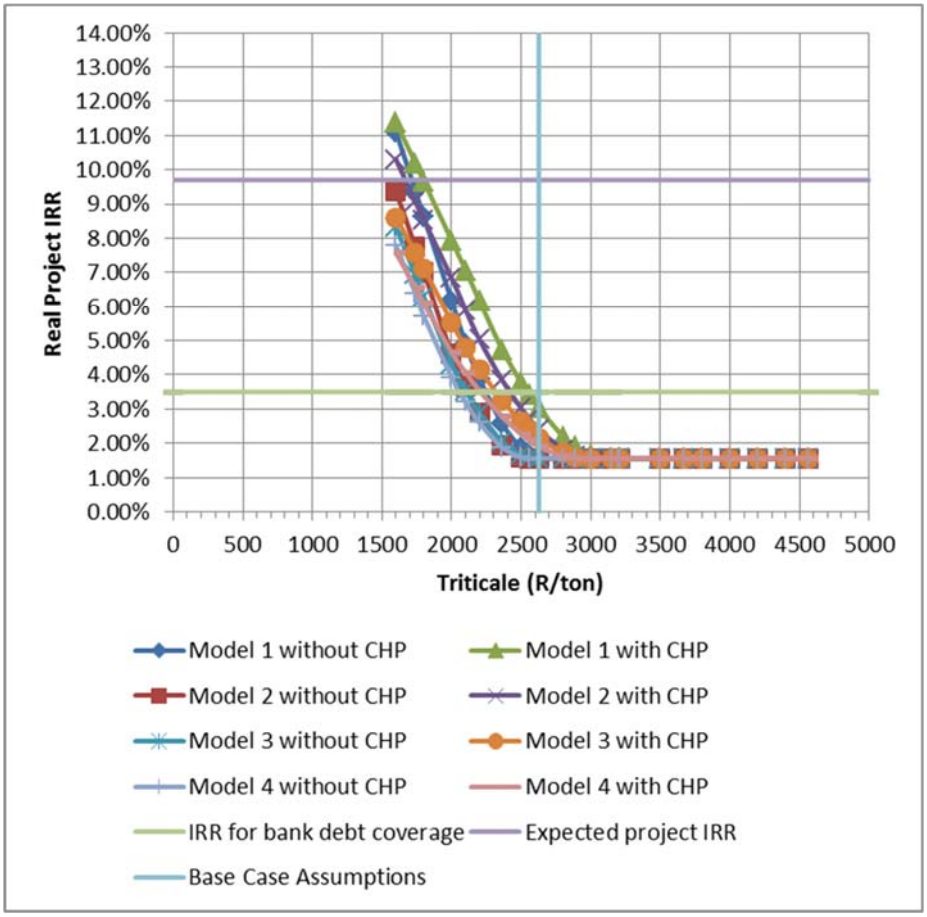

FIGURE 28: TRITICALE PRICE VERSUS IRR

Only the cold (model 2) and the warm processes with a CHP plant achieved the expected project IRR. The warm process with a CHP plant achieved the expected project IRR at a triticale price of R1795/tonne. This is the highest triticale price, given the expected project IRR, which can be absorbed while still rendering the plant profitable. The cold pre-fractionated process (model 4) without a CHP plant was the worst performing model and therefore it needed the lowest triticale price of all the models in order to still achieve the expected project IRR. The triticale price value to achieve the expected project IRR is R1410/tonne, which is much lower than the historical price variations minimum of R1739/tonne. Therefore the actual triticale price value is not likely to be achieved. The triticale price range to cover the debt payment is R2056-2543/tonne, from the worst to the best process option.

4.2.4.3 *DDGS Price*

Figure 29 shows a constant IRR before an increasing upward trend appears in the economic viability (measured as the IRR) as the price of the DDGS increased. Therefore, the profitability of the plant increased as the DDGS price increased. A difference in the IRR can be observed at the end of the DDGS ranges between the warm (model 3) and cold pre-fractionation (model 4) process and the warm (model 1) and cold (model 2) process. Here the warm and cold processes have a steeper slope than the pre-fractionated processes. The warm process with a CHP plant has the highest IRR (4.75%) and the steepest slope followed by the cold process and the warm and cold pre-fractionated processes.

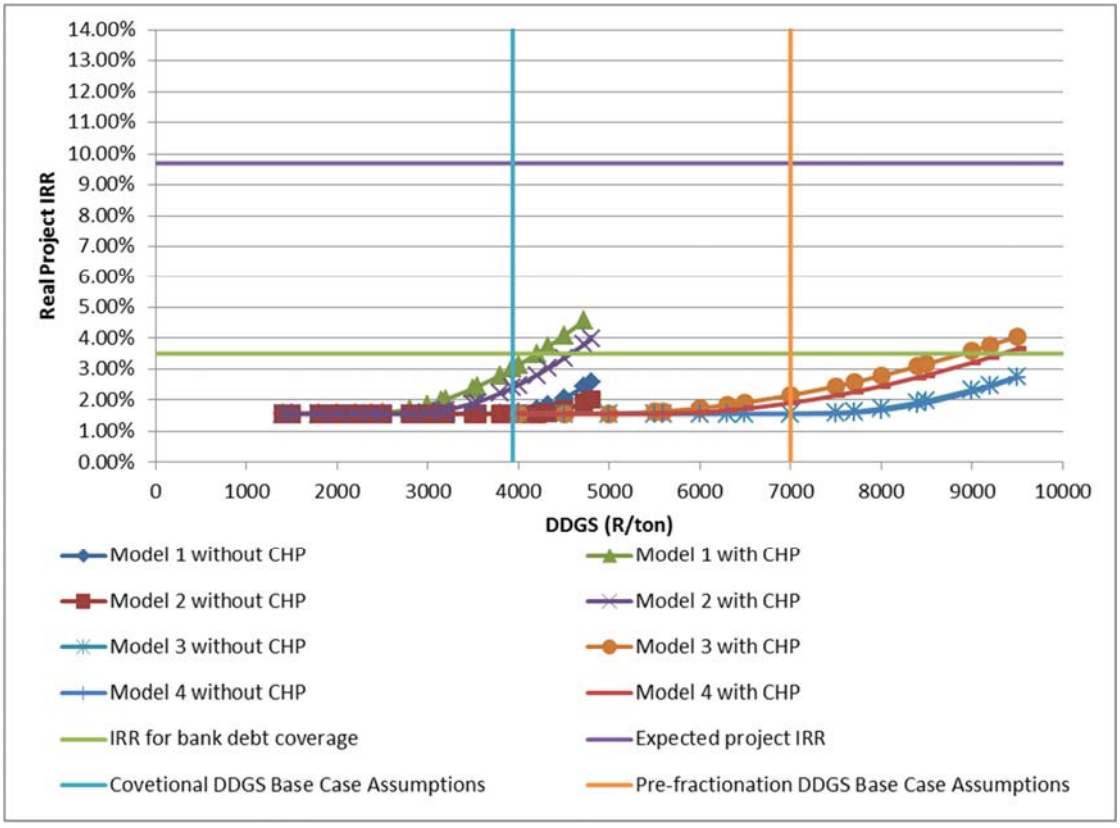

FIGURE 29: DDGS VERSUS IRR

None of the variation in the DDGS prices for any of the models could achieve the expected IRR of 9.7%. Therefore, if all other factors stay constant except the DDGS price and the DDGS price is varied within the expected range of prices, the expected project IRR cannot be achieved. A DDGS price of R6646/tonne needs to be given to the warm process with a CHP plant to achieve an IRR of 9.7%, which is outside the ranges set out for the DDGS price. There is some uncertainty surrounding the DDGS price. The sensitivity DDGS ranges that were used were therefore calculated by using all the available data obtained for this study that relates to the DDGS price.

Only the CHP plant processes are able cover the debt as shown at the end of their ranges for the warm and cold processes (between R4200-4600/tonne) and pre-fractionated processes (between R8800-9400/tonne).

4.2.4.4 *CO_2 Price*

In Figure 30 the CO_2 price has very little effect on the project IRR of any of the processes, with or without a CHP plant. The CHP plant processes have higher IRR than the processes without CHP plants. The IRR is more dependent on the process choice than the CO_2 price, with the warm process with a CHP plant outperforming all the other processes.

None of the variation in the CO_2 price range can achieve the IRR required to cover the debt payment or the expected project IRR. The warm process achieves the IRR for the debt payment at a CO_2 price of R275/tonne, which falls outside the range of CO_2 price values for the sensitivity analysis. Amigun (2012) uses a price of R132/tonne for CO_2, but there is some uncertainty as to whether CO_2 could be sold and at what price. It is therefore unlikely to achieve a price of R275/tonne for CO_2. CO_2 was also found to have only a small effect on the profitability of the process. Due to the minimal impact of CO_2 on the economic viability of the process, it is assumed that no CO_2 sales are realized.

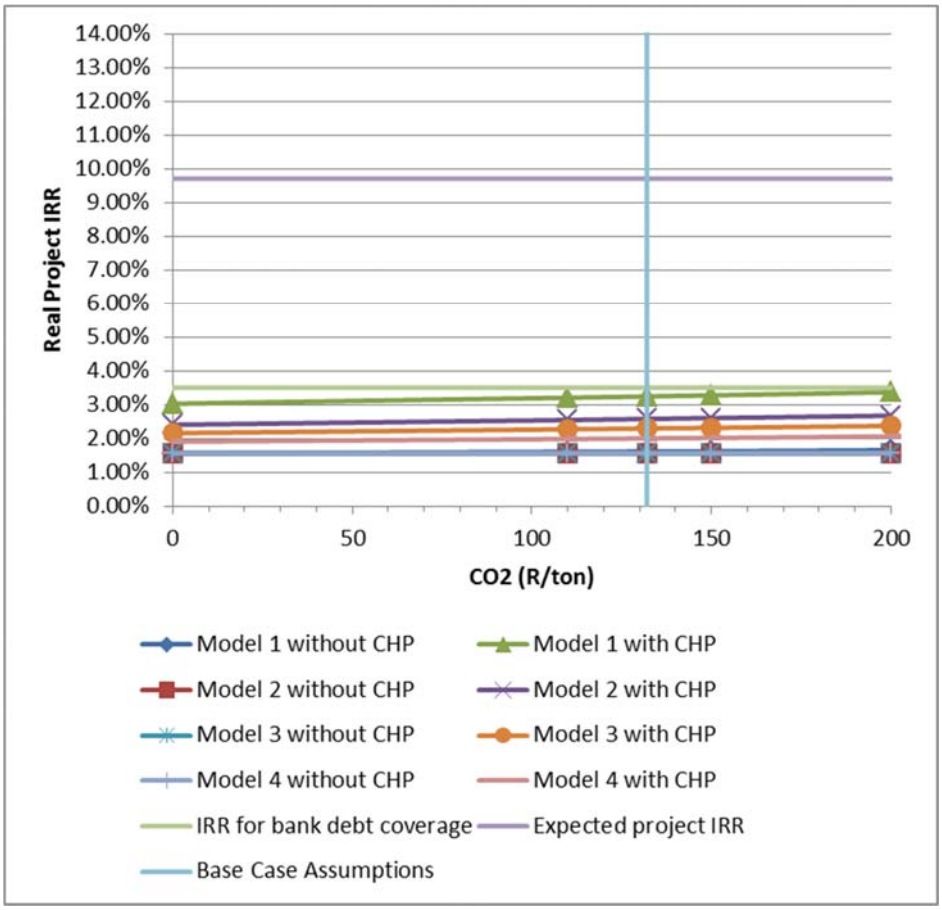

FIGURE 30: CO_2 VERSUS IRR

4.2.4.5 *Coal Price*

Figure 31 shows an overall downward trend ending in an IRR of 1.56%, as the cost of coal increases. All the CHP plant processes outperformed the processes without CHP plants. The warm process (model 1) with a CHP plant has the highest IRR followed by the cold process (model 2) with a CHP plant, the warm pre-fractionated process (model 3) with a CHP plant and finally the cold pre-fractionated process (model 4) with a CHP plant. The same successive order of processes was observed for the processes without CHP plants.

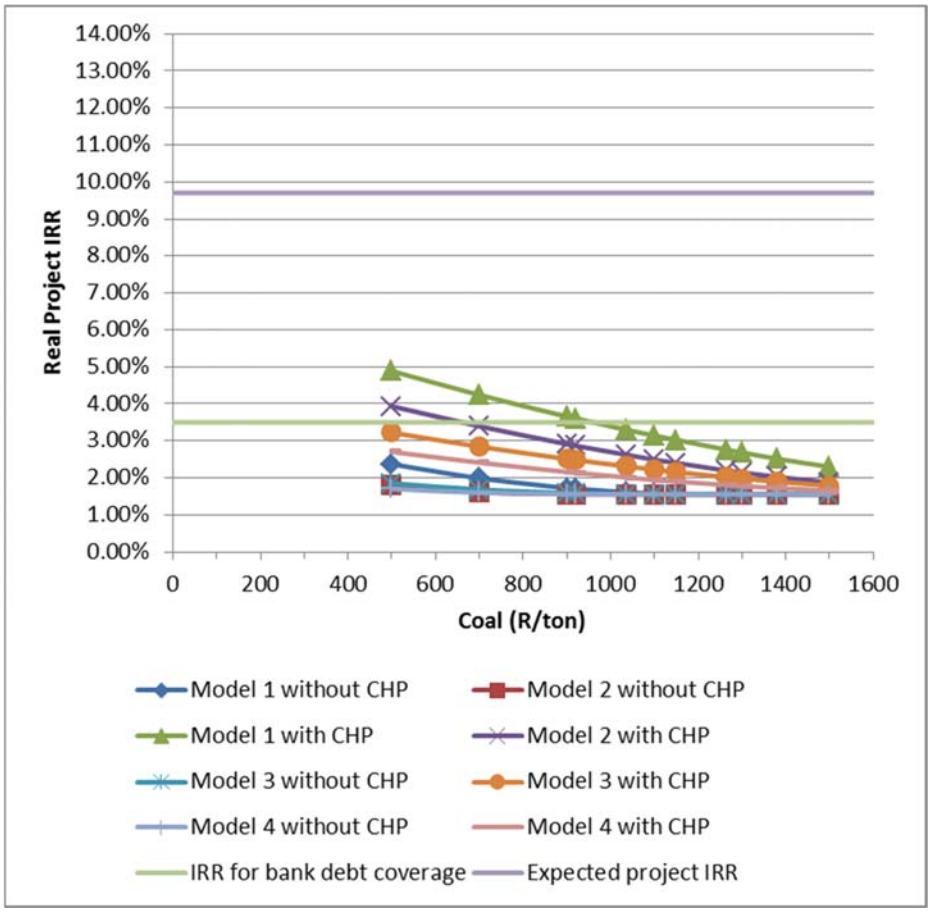

FIGURE 31: COAL VERSUS IRR

None of the processes achieve the expected project IRR threshold for the ranges considered in the sensitivity analysis. The sensitivity ranges were chosen to include all data for coal price, including buffer margins. A negative coal price would be needed to achieve the expected project IRR for all the models. Model 1 and Model 2 with a CHP plant achieves the IRR for debt coverage at a coal price of R959/tonne and R659/tonne, respectively. The worst performing model (model 4) needs a negative coal price to achieve the IRR to cover the debt. Hence, coal price does have an effect on the process profitability but to a lesser extent than the type of process. Thus the IRR is more dependent on the type of process (warm versus cold, pre-fractionation or not, and CHP versus without CHP plant) chosen than on the coal price.

4.2.4.6 *Biomass Price*

From Figure 32 it can be seen that an increase in the biomass price adversely affected the IRR of all the processes. The profitability of the warm process (model 1) with a CHP plant is the highest and therefore should be the most robust against variations in biomass price. Furthermore, the same trends as in Figure 31 are

observed except that the IRR of Figure 32 is higher than that in Figure 31.

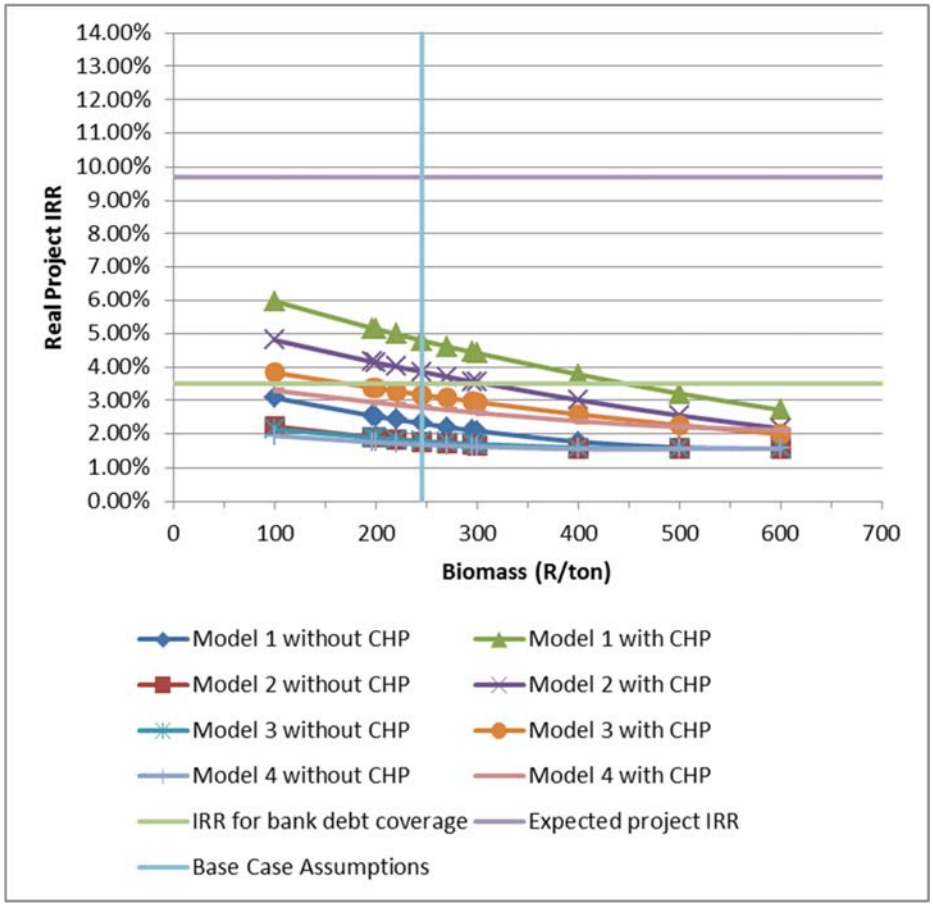

FIGURE 32: BIOMASS VERSUS IRR

The expected project IRR is not achieved by any of the processes. Negative biomass prices are needed to achieve the expected project IRR for all of the models. Three of the processes achieve the IRR to cover the debt in the sensitivity ranges given for biomass in this project. The three models are model 1 with a CHP plant, model 2 with a CHP plant and model 3 with a CHP plant, while the values required to achieve this IRR are R448/tonne, R307/tonne and R170/tonne for biomass respectively. There is also some uncertainty about the correct values for the biomass prices as it is not currently being sold in the WC. The ranges were therefore chosen to reflect the variation in the possible prices. The worst performing model, model 4, needs a negative biomass price to achieve the IRR for the debt coverage. Therefore, biomass has an effect on the profitability of the plant but the process variation and inclusion of a CHP plant have a larger effect.

4.2.4.7 *Electricity Price*

The electricity selling prices, in Figure 33, do not have a large effect on the IRR of the

process compared to BFP (Figure 26) and triticale price (Figure 28). The process that performs the best is model 1 (the warm process) with a CHP plant, as can be seen in Figure 33. The difference in the IRR between the maximum electricity selling price (R1.5/kWh) and the minimum price (R0/kWh) is 2.90% for model 1 with a CHP plant. This is not a large variation in the IRR if you take into consideration that this is the variation for the whole range of electricity selling prices. Even though the variation in IRR is small, it could still possibly affect the profitability of the process.

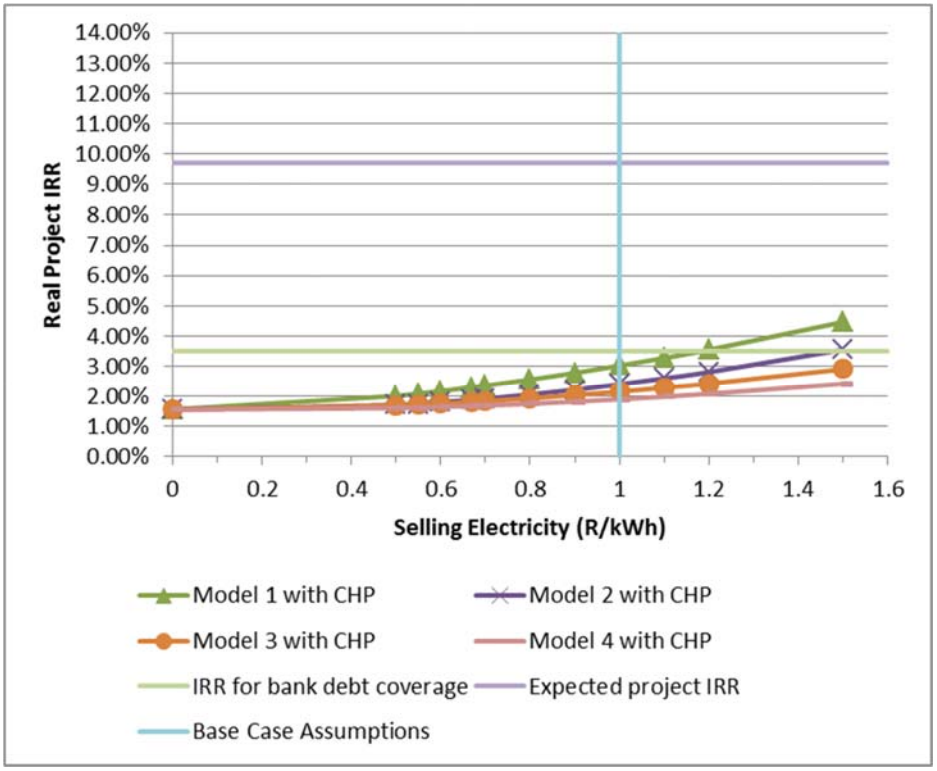

FIGURE 33: ELECTRICITY SELLING PRICE VERSUS IRR

None of the processes achieve the expected project IRR within the ranges given in the sensitivity analysis. The minimum electricity selling price for the expected project IRR was R 2.83/kWh (model 1) and the maximum was R4.87/kWh (model 4). Both of these prices are well outside the proposed range for the electricity price and therefore probably not feasible. The sensitivity ranges were chosen to cover all probable prices for selling electricity. Model 1 (the warm process) and model 2 (the cold process) achieves the IRR needed for the debt coverage in Figure 33. Model 1 achieves this IRR at an electricity selling price of R1.18/kWh and model 2 at R1.48/kWh. Model 4 achieves the IRR for the debt payment at the highest selling price of all models, which is R2.22/kWh (Figure 33).

In Figure 34 the electricity buying price has very little effect on the project IRR, as the IRR has already reached its minimum. This does not mean that the subsidy stays

constant. The subsidy increases as the buying price of electricity increases. Negative electricity buying prices are needed for all of the models to achieve the IRR required for debt coverage and the expected project IRR. Therefore, the electricity buying price under the current base case scenarios cannot render the process profitable.

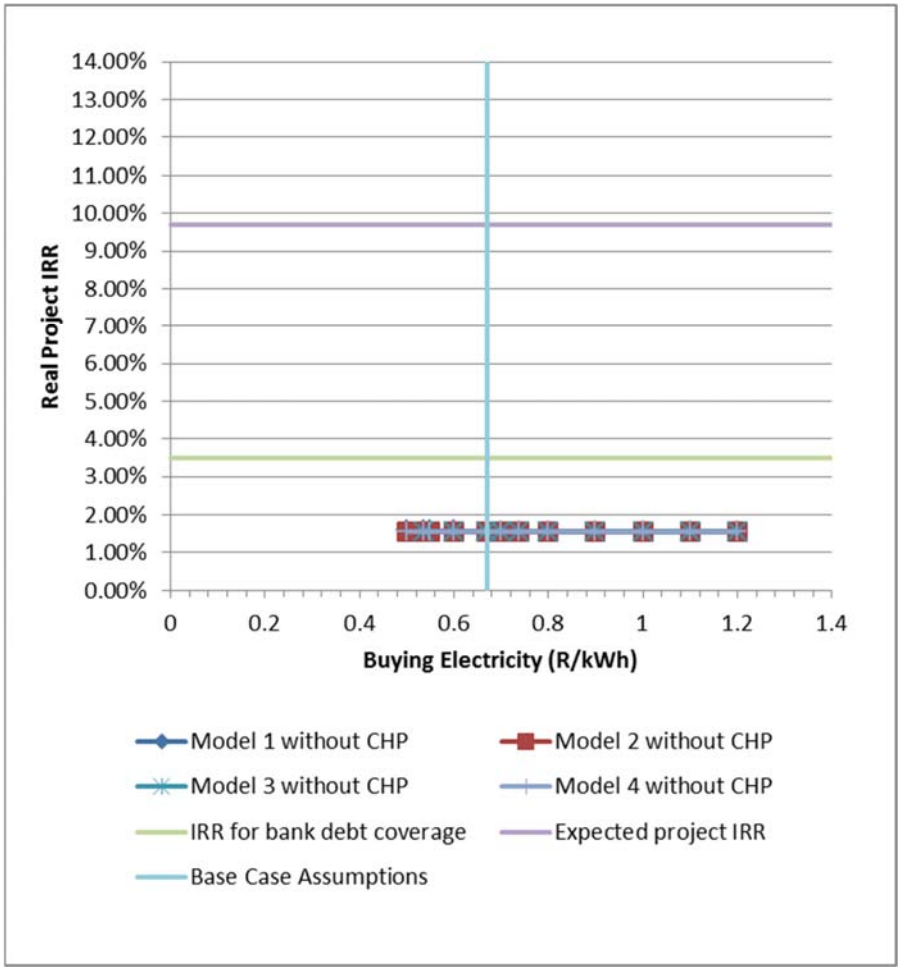

FIGURE 34: ELECTRICITY BUYING PRICE VERSUS IRR

4.2.4.8 *Production Capacity*

Figure 35 displays the impact of maintenance requirements on the production capacity (measured as the number of days annually the plant is not in operation) and consequentially on the profitability of the process. Similar trends of an increasing IRR value are observed in Figure 35, in Figure 31 and Figure 32. The warm process (model 1) with a CHP plant achieves the highest IRR. It reaches a maximum IRR of 5.29% when the plant is operational year round (no days are lost to maintenance), and a minimum IRR of 2.33% when the plant is not in operation for 50 days due to maintenance.

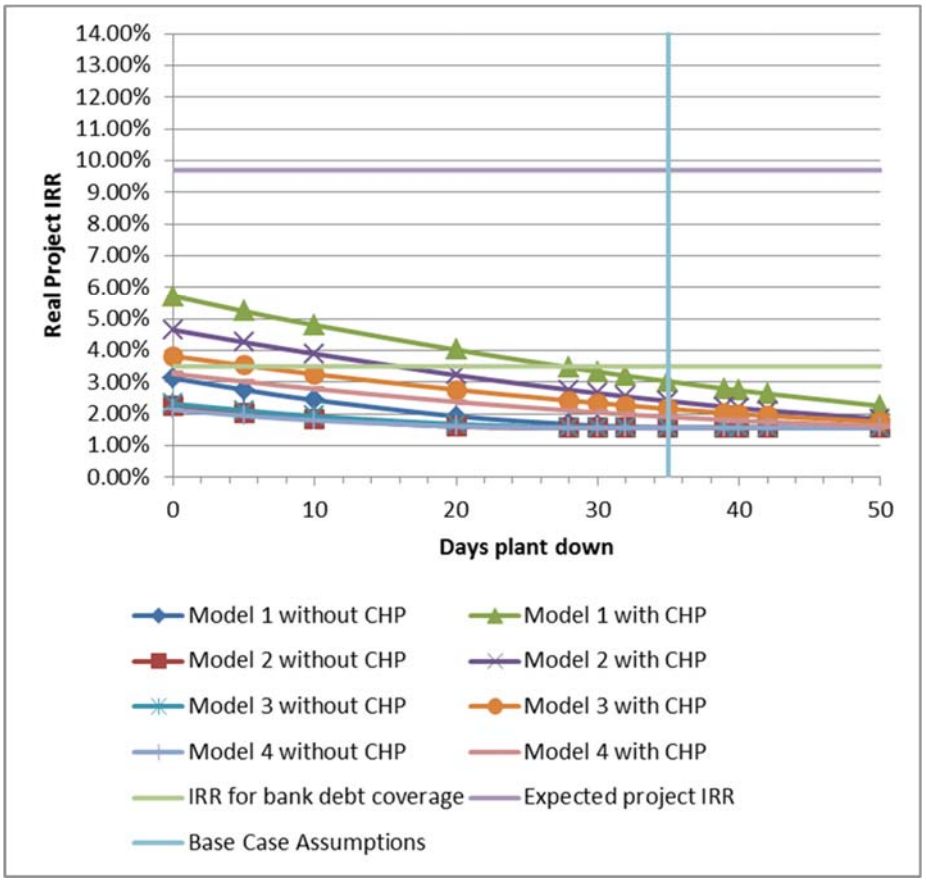

FIGURE 35: DAYS WHEN PLANT IS NOT OPERATIONAL VERSUS IRR

None of the processes achieved the expected project IRR when the number of days the plant is operational was varied. Three processes achieve the IRR required for the debt coverage. The three processes are: model 1 with a CHP plant (28 days), model 2 with a CHP plant (16 days) and model 3 with a CHP plant (6 days). The Model 1 value for the IRR required for the debt payment is obtainable as it is close to the calculated amount of days a plant is expected to be offline, which is 35 days/y (Amigun et al., 2011).

4.2.4.9 *CAPEX*

In Figure 36, the CAPEX was varied to a maximum of 20% above the calculated CAPEX and also to a minimum of 20% below the calculated CAPEX. The effect of CAPEX variation seems minor. The process that performs the best is model 1 (the warm process) with a CHP plant. It has a maximum IRR of 3.90% at 20% below the calculated CAPEX and a minimum IRR of 2.53% at 20% above the calculated CAPEX. The variation between the minimum and maximum IRR is small and therefore errors in the CAPEX calculation should not have a major effect on the IRR performance of the processes. None of the processes achieve the expected project

IRR in the ±20% range of their CAPEX. Only one process, model 1 with a CHP plant, achieves the IRR needed to make the debt payment in the ±20% range in the CAPEX value with only a 12% reduction in CAPEX being necessary. A nearly 60% reduction in the CAPEX value is needed for model 1 with a CHP plant to achieve the expected project IRR.

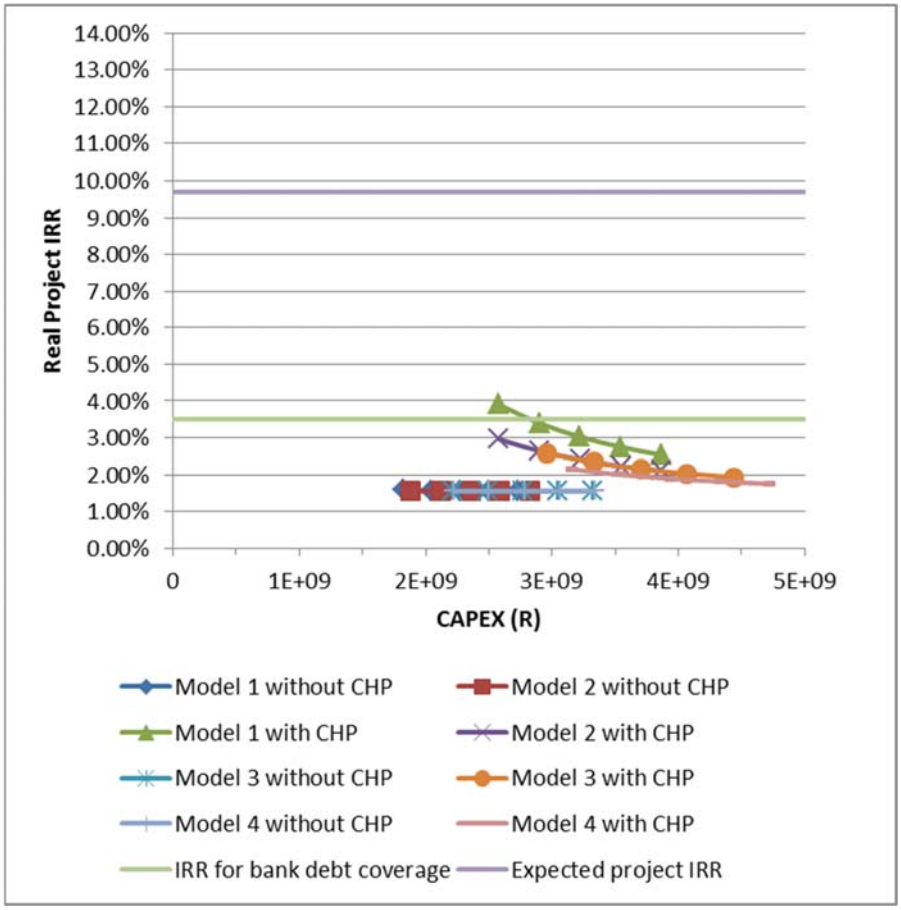

FIGURE 36: CAPEX VARIATION VERSUS IRR

4.3 Environmental Results

An analysis of the life cycle CO_2 balance for the most economically attractive process (Model 1, the warm process) was performed to compare the impact that biomass and coal have on the carbon balance.

The total amount of fossil-based CO_2 produced over the cradle-to-factory-gate section of the triticale-ethanol value chain was compared with regard to coal and biomass. This included all emissions produced in agriculture, harvesting, transport and grain conversion to reach the final ethanol product. A marked difference was observed in transport related emissions when locally sourced biomass (3 240 tonne/y) instead of coal (24 100 tonne/y) was used (see Figure 37).

Biomass leads to no CO_2 emissions being emitted by the plant as the emissions produced for steam production are carbon neutral when using biomass. This is true, since the amount of CO_2 released from burning biomass is the same amount of CO_2 consumed during the growth of the plant/feedstock (through photosynthesis) (Balat and Balat, 2009). Biomass is burned to produce heat for the process. In both cases (coal and biomass), agriculture (74 000 tonne/y) is the main contributor to CO_2 emissions.

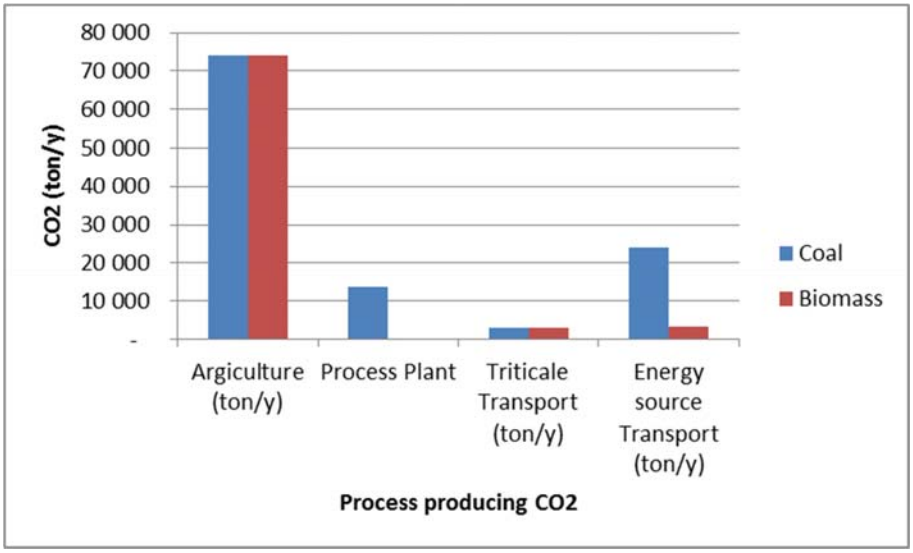

FIGURE 37: BIOMASS AND COAL CO_2 BREAK-DOWN

The following bullet points are used to explain the content represented in Table 8, row by row:

- Row 1: Shows the CO_2 emissions produced by the process using coal as its energy source;
- Row 2: Shows the CO_2 emissions produced by the process using biomass as its energy source;
- Row 3: Shows the reduction in CO_2 emissions should ethanol be used as fuel instead of petrol;
- Row 4: Shows the difference in CO_2 emissions should ethanol be used as fuel instead of petrol while using coal as fuel source. In other words, the net savings in CO_2 emissions when coal is used as the energy source;
- Row 5: Shows the difference in CO_2 emissions should ethanol be used as fuel instead of petrol while using biomass as fuel source. In other words, the net savings in CO_2 emissions when biomass is used as the energy source.

The benefit of using ethanol to replace petrol is illustrated by the reduction in CO_2 emission shown in row three of Table 8. This reduction, which represents the GHG

benefit of ethanol consumption, should be compared to the negative environmental impacts of the CO_2 emissions in ethanol production (seen in row one and two). Using biomass instead of coal significantly reduces the CO_2 emissions of cradle-to-factory-gate, due to its being carbon neutral and the smaller impact of biomass in transporting it compared to coal (see row two, Table 8). Rows 5 and 6 show the reduction in CO_2 emissions, when using bioethanol rather than petrol.

TABLE 8: CO_2 BALANCE FOR MODEL 1 WITH CHP

Models	Warm Process
Net CO2 (ton/y)	119 900
Net CO2 Biomass (ton/y)	86 500
CO2 replaced by bioethanol (ton/y)	136 100
Coal CO2 reduction (ton/y)	16 200
Biomass CO2 reduction (ton/y)	49 600

In the next section the result obtained in this chapter will be discussed.

5 DISCUSSION OF RESULTS

In this section the results will be discussed. The results compiled in the previous chapter can be divided into results pertaining to the process and results pertaining to the economics of the plant.

5.1 Process

Within this section the choice of the warm- versus cold process, pre-fractionated versus non pre-fractionated process, biomass versus coal and a CHP plant being included versus not being included in the project are discussed.

5.1.1 Warm versus cold process

The conclusion can be reached that the warm process is the preferred process, as it achieves a higher IRR (3.02%) than the cold process (2.40%); see Table 6. The difference between the warm and cold processes is that the cold process lowers the operating temperature of the enzymes used in liquefaction from 60°C to 30°C. This decrease in temperature for the cold process is negated by the use of the additional pre-saccharification step (operates at 57°C). This step is needed for improved enzyme efficiency as the yield obtained from converting starch to glucose is too low without it. Thus, the enzymes of the cold process are less efficient at converting starch to glucose than the warm process enzymes, leading to less ethanol being produced in the cold process (Sharma et al., 2009; P. Wang et al., 2007).

The cold process increases the OPEX by 2.5% as the enzymes used in this process are more expensive than in the warm process; see Table 5. The effect of the higher prices of the enzymes can be mitigated if the energy savings are significant. Since the cold process needs a pre-saccharification step and the operating temperature of this step is very close to the warm process' temperature, there is no significant energy savings.

It should furthermore also be noted that less ethanol is produced, 160 million ℓ versus 155 million ℓ, in the cold process than in the warm process. This limited ethanol production is due to the cold enzymes being less efficient when converting starch to glucose (96% versus 98%). This holds even when a pre-saccharification step is included in the cold process (see Table 3). Therefore, the cold process produces more DDGS than the warm process, due to the increased content of resistant or residual starch in the DDGS. To conclude, the IRR for the cold process is always smaller than the warm process' IRR (with less than 1%) for all different process variations (with CHP plant, biomass/coal), see Table 6. Therefore, the warm process performs marginally better and is the preferred process. This may change if the enzyme price of the cold process is lowered or the enzyme efficiency at converting starch to glucose improves in the future.

Overall the warm process out performed all of the other processes for all the different variations in economic parameters changes. The preferred process in terms of overall performance and economic robustness is the warm process and it is therefore recommended.

5.1.2 Pre-fractionated versus non Pre-fractionated Process

The selling price of the pre-fractionated DDGS should be between 2.5-4 times higher than the DDGS without pre-fractionation, to render the pre-fractionation processes more profitable than the conventional processes, see Figure 27. The pre-fractionated process removes the bran from the triticale grain prior to the final grinding for the hydrolysis-fermentation. The bran, which is high in fibre, is removed and therefore the DDGS protein content is increased. Consequently less DDGS is produced but with a higher protein content, see Table 3. This DDGS can be sold at higher prices due to it being able to be digested by non-rumen animals.

The pre-fractionation process increases the CAPEX of the plant by 12.9-17.4% due to the extra process equipment that is necessary. The higher selling price of DDGS produce should therefore mitigate the effect that the increase in the CAPEX has on the IRR (Table 5). Currently the price is estimated as only 1.8 times higher and therefore the processes without pre-fractionation outperform the pre-fractionated processes. Therefore, since the warm process IRR is less than 1.5% higher than the pre-fractionated process, the warm process is preferred above the pre-fractionated process (see Table 6). This can change if DDGS price of either the pre-fractionated process DDGS selling price increases or the conventional DDGS selling price decreases.

It should also be noted that the OPEX of the warm process is less than 1.5% (Table 5) higher than that of the pre-fractionated process and thus does not have a major effect on the profitability of the plant. The small difference in the OPEX can be contributed to a trade-off between the smaller feed streams with lower heating duty and the larger amounts of water that needs to be evaporated in the DDGS drying section. This trade-off results in energy savings occurring before the DDGS drying section which is negated by increased energy usage when the process is in the DDGS drying section (Table 4). The energy usage in the DDGS drying section increased due to less DDGS being produced for pre-fractionated processes (Table 3). Therefore, the marginal difference in energy usage caused a marginal difference in the value of the OPEX. Furthermore, the increase in enzyme usage in the process could also affect the OPEX.

5.1.3 Biomass versus coal as energy source

Biomass is the preferred energy source as it results in a larger IRR (4.79% versus 3.02% for model 1 with a CHP plant) than when coal is used. The reason for this

higher IRR is due to its smaller OPEX (see Table 6). Using Biomass as the fuel source rather than coal lowers the OPEX of the plant between 3.21% and 7.54% (Table 5). Coal needs to be transported over long distances from the northern and eastern parts of SA and therefore leads to the per unit energy supplied in the WC being more expensive than biomass.

Biomass can be harvested locally and is therefore less expensive. Furthermore, in the WC biomass (R1.1 million/MW)) has a lower market value than coal (R2.3 million/MW). It should also be noted that more biomass (2.41 x coal amount) is needed to achieve the same heating duty as coal. This is due to biomass having a lower heating value. The larger tonnages of biomass than coal used is mitigated by the price difference between biomass and coal. This price difference is large enough to alleviate the effect of the larger tonnages on the OPEX (see Table 4). It was confirmed that using biomass also lowers the carbon footprint with up to 25% (Table 8) and therefore using biomass is more environmentally friendly.

The government should encourage biomass utilization as process energy fuel. This can be done by mandating the use of a carbon neutral fuel sources for energy production in the ethanol production process. Alternatively this can be done by the government buying electricity at higher prices if it is carbon neutral or even by tax incentives. It is important that biomass be used as the preferred fuel sources as using coal can negatively influence the GHG balance. According to Wang (2007 if coal was used as energy source for steam production, this would increase the corn ethanol GHG life cycle emissions (compared to gasoline life cycle emissions) by 3% and therefore ethanol produces more GHG emissions than if gasoline was used as fuel for road transport.

5.1.4 CHP plants versus not including CHP plants

CHP plants perform better than plants with conventional boilers, since they achieve higher IRRs. The processes with CHP plants have a lower OPEX than those without CHP plants. The difference in the OPEX value can be seen in a difference of ±1% for the coal plants and between 2.37-5.38% for biomass plants (see Table 5). The small difference between the processes with CHP plants and the ones without can be contributed to the trade-off between the cost of buying electricity for the processes without CHP plants and the extra cost of buying more coal or biomass (Table 4) for the processes with CHP plants.

It should be noted that although there is only a small difference in the OPEX value, this is not true for the IRRs. The models with a CHP plant perform better despite a 25.3-29.4% increase in CAPEX (see Table 5). This better performance is due to the sale of electricity which increases the project IRR with between 0.78% and 2.11% (Table 6).

The risk corresponding to the additional CAPEX should be evaluated against the increase in IRR to determine if the CHP plant option is preferred. The necessary bidding process to acquire support from the government in the form of a subsidy and to also obtain consent from the government to produce and sell electricity should be considered. Each of these bids should be placed at the appropriate branches of the government. In addition, if the surplus electricity being exported to the national grid (±20-24MW) is produced from biomass, the electricity will be considered renewable and green which could achieve higher selling prices.

There is some uncertainty concerning the sale of electricity to the national grid by means of the Renewable Energy Independent Power Producer Procurement Programme (REIPPPP). Since, in this case if electricity sales is not realised the profitability of the plant will be negatively influenced, resulting in a decrease of the IRR to the minimum IRR value of 1.56% (see Figure 33). The government should decrease the difficulty pertaining to the selling of electricity, by making a special provision for bioethanol producers. This provision should encourage the generation of electricity in combination with bioethanol production. An additional option is to sell electricity directly to local municipalities which can simplify the selling process of surplus electricity. If all of the above points are taken into consideration, a CHP plant should be preferred to the process using the conventional boiler.

5.2 Economic Considerations

In this following section the economic results are discussed while focussing on the economic parameters and subsidy requirement of the project.

5.2.1 Economic parameters

The factor that has the largest effect on the profitability of the plant is the BFP. This can be seen by the effect a variation in the BFP has on the IRR (Figure 25). The maximum BFP from January 2009 to April 2015 was R9.21/ℓ and the minimum was R3.91/ℓ, while the IRR for the former was 11.13% and for the latter was 1.56% (see Figure 26). Therefore, the large variations in the BFP that was observed would affect the profitability of the plant. The IRR values given above relates to Model 1 with a CHP plant. This model is the model that performs the best of all the coal based models.

The BFP variability can be attributed to the volatility of the crude oil market as well as the variation in the currency exchange rates (dollar versus rand). The BFP cannot be controlled by the bioethanol producers as it is determined by internationally influenced factors. The primary way they can mitigate the effect of variations in the BFP is to fix the triticale price at a low value and fix the DDGS price at a high value. This mitigation option is discussed in detail later on in this section. The government in turn can mitigate the effect of the BFP volatility by varying the subsidy granted in

accordance to the monthly BFP changes. This is done in order for the plant to achieve economic feasibility. The government may also possibly in extreme circumstances (at a very low BFP) offer additional subsidy to the bioethanol producer to render their plants profitable. Currently the 15% ROA subsidy calculation method used by the government is insufficient to render the plant profitable; this is discussed in further detail in section 5.2.2.

Another process economic parameter that has a major effect on the profitability of the plant is the triticale price. The feedstock price is 75% of the OPEX (Figure 14) of the plant. Accordingly, the variation in the feedstock prices has a major effect on the profitability of the plant. The effect of feedstock price on the IRR can also be seen when the IRR values are considered at the maximum and minimum triticale prices from January 2009 to April 2015. The IRR values given below relates to Model 1 with a CHP plant. This model is the model that performed the best of all of the coal-based models. The IRR at the maximum triticale price (R3212/ton) was 1.56% and at the minimum price (R1739/ton) it was 11.07% (see Figure 28). The plant was found to be profitable only at a triticale price below R1800/ton given the current economic conditions (Figure 28). Therefore, the SAFEX average price of B4 wheat was found to be too high to render the plant profitable. If the triticale price is not fixed (as previously mentioned above), the subsidy calculation should be dynamic and be revised monthly in accordance to BFP and the triticale price. Under the current ROA subsidy methodology this variation is not enough to render the process profitable and therefore other options should also be investigated as is discussed later in section 5.2.2

There is some uncertainty concerning the triticale feedstock price (assumed to be the B3 or B4 wheat price), as it is not currently sold commercially. This could have a major effect on the profitability of the plant. Triticale is currently utilised on farms as animal feed and therefore its price can possibly be related to that of yellow maize, which is classified as an animal feed (R1400/ton). The Department of Agriculture in the WC estimated the price of triticale at R2000/ton if marginal lands are used for triticale production. If triticale is produce on marginal lands it should not have an influence on food production and thus there is no conflict between the production of food and fuel.

There is also some uncertainty as to whether or not the triticale price is dependent on the BFP, which can influence the ranges for a sensitivity analysis. Figure 17 illustrates a very weak positive correlation between the BFP and triticale price. This indicates that the BFP does have an influence, although small, on the triticale price, as can be seen by the fact that grain's production cost increases as the BFP increases. A detailed market analysis should be done in the future to determine an appropriate triticale price. The BFP is regulated by the crude oil price and therefore it cannot be changed, while the triticale price might be negotiable. Therefore, to increase the plant's profitability it is very important to negotiate a low feedstock price.

The DDGS price does also have an effect on the profitability of the plant, but not to the extent that both the BFP and triticale prices does. This can be seen in Figure 25 which shows the differences in IRR variations for a ±20% price variation in economic parameters. The resulting variation of the DDGS is only 3%, while that for both the BFP and triticale price is close to 6%. There is also some uncertainty concerning the DDGS price as it is not currently produced for sale. According to previous estimates made, it should trade at nearly the same price as yellow maize, as it is an animal feed. Other estimates indicate that it will trade as 58% of the soya oilcake price. These estimates were derived from a consultation with representatives of the (ruminant) animal feed industry. The estimate of the DDGS price in the pre-fractionation process is between R7000/ton and 7500/ton, which is also based on the assumption that the pre-fractionated DDGS will trade at the same price as soya oilcake. All of these factors can influence the process profitability (Figure 27) and therefore it is recommended that a detailed market analysis be done to determine the price of DDGS. Furthermore, DDGS prices should also be fixed at the highest obtainable value in contracts to render the plant profitable.

The other process parameters namely coal, biomass, CO_2, CAPEX, capacity and electricity selling and buying price have less than a 1% effect on the profitability of the plant (see Figure 25). CO_2 has almost no effect, as can be seen in Figure 30. Coal (Figure 31) and Biomass (Figure 32) also do not have a large effect on the IRR of the process. The IRR is more dependent on the type of process chosen than on the process parameters. This holds for all of the parameters except for the BFP and triticale price parameters. Therefore, given that the BFP and triticale price is favourable, the type of processes chosen becomes the biggest influential factor. From all of the types of processes that were considered for this project, the warm process with a CHP plant was found to be the preferred process.

Also of interest is that there is interplay between the BFP and triticale price and the inclusion of a CHP plant. At low product (BFP) / high feedstock (triticale) prices, the process with a CHP plant performs better than those without CHP plants. At higher product and lower feedstock prices, i.e. more favourable condition, the type of process used determines how well the process will perform (see Figure 26, Figure 28 and Figure 18). If the conditions are unfavourable, all of the processes with CHP plants outperform those without CHP plants. At more favourable conditions the best performing project plant is more process dependent (e.g. cold versus warm process). Therefore there is a trade-off between processes at different economic conditions.

5.2.2 Subsidy

It is possible to reach zero subsidy (only under favourable conditions), but the IRR achieved under this condition is still very low. Zero subsidy is therefore not attractive to investors except if a high enough IRR can be achieved under the process conditions required to obtain a zero subsidy; see section 4.2.2 and Figure 22. Hence

financial support is needed from the government to render the process profitable.

The current subsidy is calculated on 15% ROA and in most cases project plants do not achieve the desired project IRR. By changing the way the assets are depreciated for the ROA calculation, the 15% ROA calculation can be manipulated, and therefore there are different ways to calculate the subsidy. There is a relationship between the IRR and the ROA. The minimum IRR will therefore be reached at the subsidy value (R2/ℓ) after which the IRR stays constant even though the subsidy value continues to increase. When straight line depreciation of assets method the project IRR's minimum was calculated as 1.56%; see Figure 22. These IRR base values were observed to be below the project's real IRR value (9.7%), which the project needs to achieve to be profitable.

Consequently, it is recommended that alternative subsidy calculations should be investigated. A possible way to calculate the subsidy value that could be investigated is done by specifying the IRR value and then variating the value of the subsidy until the specified IRR is reached. The subsidy value needed to obtain the specified IRR is the required subsidy value. Furthermore another way to increase the subsidy amounts are by adjusting the % ROA used for the subsidy calculation from 15% to 20%. It should also be noted that a clawback mechanism is not currently applied in the models under discussion in this project.

If the government uses sorghum as its reference grain feedstock for subsidy purposes while sorghum is more expensive than triticale, then triticale has an economic advantage**.** Sorghum should be more expensive than triticale as triticale is produced locally and the transport differential that applies to sorghum does not apply to triticale. If it should be the case that sorghum is cheaper than triticale, the subsidy will be insufficient and a negative IRR (Figure 23) will result. Therefore, given the uncertainty surrounding the triticale price, sorghum can have either a large positive or a large negative effect on the IRR. Therefore the bioethanol producers in the WC (that uses triticale grain) experience greater financial risk if sorghum is used as the reference grain, than bioethanol producer that uses sorghum. As sorghum and triticale are grown in different parts of SA, their yields can differ depending on environmental factors. It can be argued that the process energy is more expensive if coal is used in the WC rather than in the northern parts of SA. Triticale bioethanol producers should therefore receive a larger subsidy, or triticale grain prices should be used as their reference grain for their plants. The effect of the triticale price variation in relation to the sorghum price should be taken into consideration in the subsidy calculation mechanism for triticale ethanol plants.

It should furthermore be noted that the results obtained in the project are based on a theoretical simulation and assumptions made in regards to the process parameters, utilities and feedstock prices. The process parameters need to be confirmed by experimental work and pilot plant testing, while price assumptions need to be

evaluated through the use of more detailed economic analysis. During the process the simulated models could be updated as more information becomes available, in order to assess and enumerate the overall technical and economic feasibility of the models on an ongoing basis.

In the next section the conclusions and recommendations drawn from this section are presented.

6 CONCLUSION AND RECOMMENDATIONS

From the previous section the following conclusions can be drawn and the following recommendations are made.

- Model 1 (the warm process) produces the most ethanol from triticale, which is 160 million ℓ/year. This is due to the enzymes in the warm process being more effective than those in the cold process, as well as the fact that the bran is not removed from the grain for the warm process as in the pre-fractionated process. In the pre-fractionation processes some starch is lost when the bran is removed, as the triticale hull is not a smooth but rather a wrinkled surface.
- Model 1 (the warm process) without a CHP plant achieved the lowest CAPEX (R2.274 billion).
- Model 4 (the cold pre-fractionated process) with a CHP plant has the highest CAPEX (R3.904 billion). Pre-fractionation requires extra process equipment to remove the bran from the triticale grain and therefore the CAPEX increases. The cold process is furthermore less efficient in the conversion of starch to glucose due to enzyme inefficiency and therefore larger amounts of DDGS are produced. This leads to an increase in the size of equipment that is needed.
- Due to enzyme inefficiency at converting starch to glucose during the cold process, the warm process achieved a higher IRR. It is therefore recommended that the warm process be preferred above the cold process.
- Model 4 has the lowest heating duty at 79.4MW. This is due to the dry pre-fractionation process, which reduces the amount of DDGS than needs to be dried, and the use of an enzyme that needs a lower operating temperature during liquefaction.
- Model 1 has the highest heating duty at 86.3MW. This is due to the warm process being the most energy intensive process as it has no specified energy saving process altercations except for heat integration. Thus, it does not operate at a lower temperature for liquefaction and its feed stream is larger than those in the pre-fractionated processes are. It therefore consumes more energy in the process before the DDGS drying section than the other processes.
- Using Biomass as fuel source rather than coal lowers the OPEX of the plant by between 3.21% and 7.54% and therefore the IRR of the process increases (0.13-1.77%). This lower OPEX is due to coal (R2.3 million/MJ) being more expensive in the WC, since it is transported over long distances from the northern and eastern parts of SA. Biomass (R1.1 million/MW) can be harvested locally and therefore it need not be transported over great distances.
- Using biomass to replace coal as energy source significantly lowers (25%) the CO_2 emissions of the plant. Hence, it is recommended that biomass should be used as the plant's energy source as it is carbon neutral (more environmentally friendly) and less expensive.

- It is recommended that the government strongly encourage the usage of biomass for process heat. This can be done by either (1) mandating a carbon neutral fuel source for energy needs, (2) buying electricity at a higher price from the ethanol producer or even (3) offering tax incentives.
- The models perform better with a CHP plant despite the resulting higher CAPEX (25.3%-29.4% increase in CAPEX). This can be attributed to the selling of electricity, which increases the project's IRR by between 0.78% and 2.11%. The conclusion can be drawn that the inclusion of a CHP plant to the process is the preferred option.
- The inclusion of CHP plants results in lower OPEX values than those processes without CHP plants. The OPEX decreases ±1% for the coal plants and between 2.37% and 5.38% for the biomass plants. This difference can be assigned to the trade-off between the extra coal and electricity that was bought.
- It is recommended that the government simplify the process for selling the surplus electricity obtained during the bioethanol production process, by introducing a special provision for biofuels producers.
- Another option to simplify the selling of surplus electricity is to sell it directly to local municipalities.
- Model 1 is the model with the highest electricity production (34.3MW), due to it having the highest heating duty. If the amount of steam is increased, the amount of electricity produced will also increase. In other words, there is a positive correlation between the amount of steam and electricity produced.
- The BFP and triticale price are the two factors that have the most dominant effects on the profitability of the plant. This can be seen as the IRR varies from 1.56% to 11.13% in correspondence with a variation in the BFP from the maximum BFP (R9.21/ℓ) to the minimum BFP (R3.91/ℓ). The IRR varies from 1.56% to 11.07% when the triticale price is varied between the maximum (R3212/ton) and minimum (R1739/ton) triticale price. Therefore, the subsidy calculations should be dynamic in addition to other subsidy alternations being made to render the process profitable. The subsidy calculation can be made dynamic by revising the subsidy monthly in accordance with the BFP and triticale price.
- There is uncertainty in regard to the triticale and DDGS prices and therefore a detailed market assessment is recommend to determine these prices
- Triticale price should be fixed below R1800/ton in a contract to render the process profitable, to reduce the effect of the BFP on the profitability of the process. This price is well below the current SAFEX price for B4 wheat.
- DDGS price should be fixed as high as possible to render the process profitable.
- If triticale is used for ethanol production which is produced on marginal lands it will not influence food production.

- Due to the increase in the CAPEX (13-17%) associated with the pre-fractionated processes (extra process equipment required for the removal of bran), the warm process performs better than both of the pre-fractionation processes do. This increase in CAPEX must be justified by a significant increase of at least 2.5-fold in the DDGS selling price, since it can be sold as non-rumen animals feed. Currently the difference is only 1.8 times, which is too low to render the warm pre-fractionation process more profitable than the warm process. Given the current process conditions the warm process was found to be the most robust process for all the sensitivity analyses. This is only true if the DDGS of the pre-fractionated process is not more than 2.5 times higher than the warm process' DDGS. If it is however higher, the warm pre-fractionated process would be the most robust process. Therefore, the warm process is recommended above both of the pre-fractionated processes under the current economic conditions.
- CO_2 has almost no effect (less than 0.5%) on the IRR of the plant and therefore it is not a major contributing factor to the profitability of the plant.
- The effect of price variation in the price of coal, biomass and electricity buying and selling prices as well as in the value of CAPEX and the plant's production capacity has a small effect (±1%) on the IRRs of the different plants that were modelled.
- Zero subsidy can be achieved for model 1 with a CHP plant at a BFP of R7.94/ℓ and a triticale price of R2578/ton, or at a triticale price of R2232/ton and a BFP price of R6.94/ℓ. Therefore it is possible to achieve zero subsidy, but this will only happen under favourable conditions and is not the norm. It should be noted that the IRR achieved at zero subsidy is 5.85%, which is well below the expected project IRR of 9.7%. The return on investment will therefore be lower than desired.
- Using sorghum as a references grain for the subsidy calculation can have a large positive or negative effect on the profitability of the plant. The profitability of the plant in turn is subject to the price of sorghum in relation to the price of triticale. The effect of the triticale price variation in relation to the sorghum price should be taken into consideration for the subsidy calculation mechanism used for triticale ethanol plants. This should be done to make the WC (triticale) plants more competitive compared to other ethanol plants in SA.
- The fixed asset value ROA method results in a minimum IRR of 8.09%, that is higher than the minimum IRR (1.56%) obtained by means of the straight line depreciated asset value ROA method. This is due to the fact that the fixed asset value method always calculates the theoretical maximum subsidy that corresponds to the current 15% ROA.
- The model with the highest IRR with a selling price of triticale of R2578/tonne and DDGS of R3935/tonne is model 1 with a CHP plant. It has an IRR of 3.02% for the straight line ROA method. It should also be noted that this value is below the IRR needed to cover debt repayment (3.5%) and therefore it is not economically

feasible. It is therefore recommended that an alternative subsidy calculation method should be investigated to render the plant profitable.

- The process that is recommended as the process to use for the production of DDGS and ethanol from triticale, is the warm process with a CHP plant. This recommendation is made as this model has the highest IRR and is well established internationally.
- Currently there exist uncertainty as to the true price of triticale and DDGS. If alternative prices are therefore taken into account (R2000/ton for triticale and R1800/ton for DDGS) the model with the highest IRR will be model 3 (the warm pre-fractionation process) with a CHP plant. It has an IRR of 5.53%. Hence the IRR is highly dependent on the triticale and DDGS prices. A thorough investigation into these prices by means of a market analysis is therefore recommended.
- The IRR can be used to determine the subsidy amount and not ROA which would lead to a more profitable plant
- Another way to improve the profitability of the plant is to increase the ROA subsidy from 15% to 20%.
- Additional in when the BFP is very low their can maybe be additional subsidy support to render the plant more profitable.
- It should be noted that the results are based on theoretical simulations and assumptions regarding processes, parameters, utilities and feedstock prices. Furthermore, process parameters need to be confirmed by experimental work and pilot plant testing before actively using them, while price assumptions need to be evaluated by more detailed economic studies. During the process, the simulated models should be updated as information becomes available to assess and evaluate the overall technical and economic feasibility on an ongoing basis.

The following further research is recommended:

- Evaluate at which economic scale (capacity) bioethanol production from triticale becomes economically feasible;
- Evaluate whether a higher protein or a higher starch content in triticale is more desirable for economic feasibility;
- Evaluate how much yeast producing enzymes have an effect on the project's profitability;
- Evaluate for the cold process if an increased fermentation time is more economically feasible than a shorter fermentation times due to an increased enzyme dosage;
- Evaluate if the cold process could be done without the pre-saccharification and if so, what effect this would have on the profitability of the plant;

- Do pilot plant and laboratory testing, using locally produced triticale, of the models created in this thesis;
- Do further modelling on the use of biomass boilers and CHP plants in a triticale ethanol plant to investigate if coal and biomass has the same CAPEX and performance here; and
- Do complete Life Cycle Assessments (LCA) for all of the processes to evaluate their environmental impacts.

Concluding remarks

In this project the assessment of various processes for triticale fermentation to ethanol and DDGS were modelled and designed for the WC province. In conclusion, this project recommends the warm process with a CHP plant using biomass as fuel for its energy source for ethanol production from triticale. Furthermore, it is recommended that an investigation, that includes a detailed market analysis, should be done to obtain accurate fixed values for both the triticale and DDGS price. These fixed prices should be used in DDGS and triticale contracts. Under current economic conditions the available subsidy is insufficient to render the plant profitable. Consequently, alternative subsidy calculation variations and mechanisms should be investigated. In addition, it is also recommended that further testing on all of the considered processes should be done on a lab and pilot plant scale.

7 REFERENCES

Abebe, M., 2008. History of ethanol.

Adams, J.M.M., Teunissen, P.J.M., Robson, G., Dunn-Coleman, N., Donnison, I.S., 2012. Scanning electron microscopy and fermentation studies on selected known maize starch mutants using STARGEN™ enzyme blends. Bioenergy Res. 5, 330–340.

Amigun, B., Petrie, D., Görgens, J., 2012. Feedstock and Technology Options for Bioethanol Production in South Africa: Technoeconomic Prefeasibility Study. Energy Fuels 26, 5887–5896. doi:10.1021/ef3008272

Amigun, B., Petrie, D., Görgens, J., 2011. Economic risk assessment of advanced process technologies for bioethanol production in South Africa: Monte Carlo analysis. Renew. Energy 36, 3178–3186. doi:10.1016/j.renene.2011.03.015

Balat, M., Balat, H., 2009. Recent trends in global production and utilization of bio-ethanol fuel. Appl. Energy 86, 2273–2282. doi:10.1016/j.apenergy.2009.03.015

Balat, M., Balat, H., Öz, C., 2008. Progress in bioethanol processing. Prog. Energy Combust. Sci. 34, 551–573. doi:10.1016/j.pecs.2007.11.001

Balcerek, M., Pielech-Przybylska, K., 2013. Effect of simultaneous saccharification and fermentation conditions of native triticale starch on the dynamics and efficiency of process and composition of the distillates obtained: simultaneous saccharification and fermentation of native triticale starch. J. Chem. Technol. Biotechnol. 88, 615–622. doi:10.1002/jctb.3873

Best, M.J., Shumaker, G.A., Flanders, A., McKissick, J.C., 2005. The economic feasibility of operating an advanced ethanol production facility in Georgia.

Biology-Online, 2015. Biology-Online Dictionary [WWW Document]. URL http://www.biology-online.org/dictionary?sid=e6b1c4427fbb831cf6d7b4f8306f6ae8 (accessed 2.21.16).

Biomass Energy Centre, 2011. Typical calorific values of fuels [WWW Document]. URL http://www.biomassenergycentre.org.uk/portal/page?_pageid=75,20041&_dad=portal&_schema=PORTAL (accessed 1.23.15).

Bothast, R.J., Schlicher, M.A., 2005. Biotechnological processes for conversion of corn into ethanol. Appl. Microbiol. Biotechnol. 67, 19–25. doi:10.1007/s00253-004-1819-8

Boundy, B., Diegel, S.W., Wright, L., Davis, S.C., 2011. Biomass Energy data Book.

Bowonder, B., 1983. The alcohol economy: Fuel ethanol and the Brazilian experience: by H. Rothman, R. Greenshields and F. Rosillo CalléFrances Pinter, London, 1983, 198 pp,\pounds 13.75. Elsevier.

Bureau of Transportation Statistics, 2012. National Transportation Statistics. USA.

Cai, H., Dunn, J.B., Wang, Z., Han, J., Wang, M.Q., others, 2013. Life-cycle energy use and greenhouse gas emissions of production of bioethanol from sorghum in the United States. Biotechnol Biofuels 6, 1–15.

Cardona, C.A., Sánchez, Ó.J., 2007. Fuel ethanol production: Process design trends and integration opportunities. Bioresour. Technol. 98, 2415–2457. doi:10.1016/j.biortech.2007.01.002

Cilliers, T., 2015. Stargen Price.

Coetzee, L., 2015. B3 Wheat pricing.

Danisco US Inc, 2009. STARGEN™ 002: Granular Starch Hydrolyzing Enzyme for Ethanol Production.

Demirbas, A., 2008. The Importance of Bioethanol and Biodiesel from Biomass. Energy Sources Part B Econ. Plan. Policy 3, 177–185. doi:10.1080/15567240600815117

Department of Energy, 2014a. POSITION PAPER ON THE SOUTH AFRICAN BIOFUELS REGULATORY FRAMEWORK "The First Phase of the Biofuels Industrial Strategy." Pretoria.

Department of Energy, 2014b. Petroleum | Department: Energy | REPUBLIC OF SOUTH AFRICA [WWW Document]. URL http://www.energy.gov.za/files/esources/petroleum/petroleum_arch.html (accessed 4.7.14).

Department of Minerals and Energy, 2007. Biofuels Industrial Strategy of the Republic of South Africa.

Dhavala, P., Goswami, A., Maithel, S., 2006. Liquid Biofuels in South Asia: Resources and T ces and T ces and Technologies echnologies. Nat. Gas 34, 1.

Dias, M.O.S., Modesto, M., Ensinas, A.V., Nebra, S.A., Filho, R.M., Rossell, C.E.V., 2011. Improving bioethanol production from sugarcane: evaluation of distillation, thermal integration and cogeneration systems. Energy 36, 3691–3703. doi:10.1016/j.energy.2010.09.024

Eskom, 2015. Eskom Tariffs and Charges 2015/2016.

Ethanolhistory, 2011. Ethanol History - From Alcohol to Car Fuel [WWW Document]. URL http://www.ethanolhistory.com/ (accessed 10.30.15).

Federal Trade Commission, 2012. Paying a Premium for High Octane Gasoline? | Consumer Information [WWW Document]. URL http://www.consumer.ftc.gov/articles/0210-paying-premium-high-octane-gasoline (accessed 10.13.15).

Gago, F., Horváthová, V., Ondáš, V., Šturdík, E., 2013. Assessment of waxy and non-waxy corn and wheat cultivars as starch substrates for ethanol fermentation. Chem. Pap. 68, 300–307. doi:10.2478/s11696-013-0454-1

García-Aparicio, M., Trollope, K., Tyhoda, L., Diedericks, D., Görgens, J., 2011. Evaluation of triticale bran as raw material for bioethanol production. Fuel 90, 1638–1644. doi:10.1016/j.fuel.2010.10.049

Google maps, 2015. Saldanha Bay to Swellendam, N2 National, Swellendam - Google Maps [WWW Document]. URL https://www.google.co.za/maps/dir/Saldanha+Bay/Swellendam,+N2+National,+Swellendam/@-33.6322521,18.0915577,8z/data=!3m1!4b1!4m14!4m13!1m5!1m1!1s0x1c34a9293b9d5743:0x708f29ae23de67c!2m2!1d17.9887402!2d-33.0389488!1m5!1m1!1s0x1dd3d199572bb737:0x618f7d4bcecae230!2m2!1d20.44444!2d-34.02944!3e0?hl=en (accessed 11.13.15).

Govindaswamy, S., Vane, L.M., 2007. Kinetics of growth and ethanol production on different carbon substrates using genetically engineered xylose-fermenting yeast. Bioresour. Technol. 98, 677–685. doi:10.1016/j.biortech.2006.02.012

Hicks, K.B., Flores, R.A., Taylor, F., McAloon, A.J., Moreau, R.A., Johnston, D.B., Senske, G.E., Brooks, W.S., Griffey, C.A., 2005. Current and Potential Use of Barley in Fuel Ethanol Production. Presented at the 2005 EWW/SSGW Conference.

Humbird, D., Davis, R., Tao, L., Kinchin, C., Hsu, D., Aden, A., Schoen, P., Lukas, J., Olthof, B., Worley, M., others, 2011. Process design and economics for biochemical conversion of lignocellulosic biomass to ethanol: dilute-acid pretreatment and enzymatic hydrolysis of corn stover.

ICM Inc., 2012. ICM INC - Ethanol Production Process [WWW Document]. URL http://www.icminc.com/innovation/ethanol/ethanol-production-process.html (accessed 10.12.15).

Immik, H., 2008. CARBON FOOTPRINT AND CARBON NEUTRAL CALCULATIONS FOR Backsberg Estate.

Kaliyan, N., Morey, R.V., Tiffany, D.G., 2011. Reducing life cycle greenhouse gas emissions of corn ethanol by integrating biomass to produce heat and power at ethanol plants. Biomass Bioenergy 35, 1103–1113. doi:10.1016/j.biombioe.2010.11.035

Karhammar, R., Sanghvi, A., Fernstrom, E., Aissa, M., Arthur, J., Tulloch, J., Davies, I., Bergman, S., Mathur, S., 2006. Sub-Saharan Africa: introducing low-cost methods in electricity distribution networks. Energy Sector Management Assistance Program.

Kasavi, C., Finore, I., Lama, L., Nicolaus, B., Oliver, S.G., Toksoy Oner, E., Kirdar, B., 2012. Evaluation of industrial Saccharomyces cerevisiae strains for ethanol production from biomass. Biomass Bioenergy 45, 230–238. doi:10.1016/j.biombioe.2012.06.013

Kučerová, J., 2007. The Effect of Year, Site and Variety on the Quality Characteristics and Bioethanol Yield of Winter Triticale. J. Inst. Brew 2, 142–146.

Kwiatkowski, J.R., McAloon, A.J., Taylor, F., Johnston, D.B., 2006. Modeling the process and costs of fuel ethanol production by the corn dry-grind process. Ind. Crops Prod. 23, 288–296.

Lallemand, 2015. Fuel Ethanol | Lallemand Biofuels & Distilled Spirits [WWW Document]. URL http://www.lallemandbds.com/fuel-ethanol/ (accessed 11.18.15).

LD Jellison, Inc., 2011. Biomass Fuel Resource Study.

Lemmer, W., Schoeman, B., 2011. An Assessment of the Food Security Impact in South Africa and the World due to the South African Biofuels Industry Rollout. Grain SA, Bothaville.

Li, J., Vasanthan, T., Bressler, D.C., 2012. Improved cold starch hydrolysis with urea addition and heat treatment at subgelatinization temperature. Carbohydr. Polym. 87, 1649–1656.

Lin, T., Rodríguez, L.F., Eckhoff, S.R., 2011a. Process modeling of the Quick-germ/Quick-fiber process: Energy, water, and economics. Ind. Crops Prod. 34, 986–993. doi:10.1016/j.indcrop.2011.03.003

Lin, T., Rodríguez, L.F., Li, C., Eckhoff, S.R., 2011b. An engineering and economic evaluation of wet and dry pre-fractionation processes for dry-grind ethanol facilities. Bioresour. Technol. 102, 9013–9019. doi:10.1016/j.biortech.2011.06.013

MacLean, H.L., Lave, L.B., 2003. Evaluating automobile fuel/propulsion system technologies. Prog. Energy Combust. Sci. 29, 1–69.

Mashoko, L., Mbohwa, C., Thomas, V.M., 2013. Life cycle inventory of electricity cogeneration from bagasse in the South African sugar industry. J. Clean. Prod. 39, 42–49. doi:10.1016/j.jclepro.2012.08.034

McAloon, A., Taylor, F., Yee, W., Ibsen, K., Wooley, R., 2000. Determining the cost of producing ethanol from corn starch and lignocellulosic feedstocks. Natl. Renew. Energy Lab. Rep.

McGrane, S.J., Mainwaring, D.E., Cornell, H.J., Rix, C.J., 2004. The Role of Hydrogen Bonding in Amylose Gelation. Starch - Stärke 56, 122–131. doi:10.1002/star.200300242

McLean, J., 2015. Stargen Price. [Personal Communication]

Melamu, R., 2015. Biofuels: from viability to pilot projects. GreenCape.

Metals Consulting International, 2015. coking coal prices metallurgical coal steelmaking costs [WWW Document]. Steelomthenet.com. URL http://www.steelonthenet.com/files/metallurgical-coal.html (accessed 11.16.15).

Meyer, F., Strauss, P., Cutts, M., de Beer, J., du Toit, G., Funke, T., Gebrehiwet, Y., 2005. Bio-ethanol production in South Africa : An objective Analysis. University of Pretoria.

Mojovic, L., Pejin, D., Grujic, O., Markov, S., Pejin, J., Rakin, M., Vukasinovic, M., Nikolic, S., Savic, D., 2009. Progress in the production of bioethanol on starch-based feedstocks. Chem. Ind. Chem. Eng. Q. 15, 211–226. doi:10.2298/CICEQ0904211M

Nasterlack, T., von Blottnitz, H., Wynberg, R., 2014. Are biofuel concerns globally relevant? Prospects for a proposed pioneer bioethanol project in South Africa. Energy Sustain. Dev. 23, 1–14. doi:10.1016/j.esd.2014.06.005

Nkomba, E., 2015. Effect of sorghum grain decortication on bioethanol production technologies. Stellenbosch: University of Stellenbosch.

Novozymes, 2016. Liquefaction [WWW Document]. URL http://www.novozymes.com/en/solutions/bioenergy/starch-based-ethanol/liquefaction/Pages/default.aspx (accessed 2.21.16).

Nowell, M.S., 2011. Determining the hydrological benefits of clearing invasive alien vegetation on the Agulhas Plain, South Africa. Stellenbosch: University of Stellenbosch.

Nsaful, F., Görgens, J.F., Knoetze, J.H., 2013. Comparison of combustion and pyrolysis for energy generation in a sugarcane mill. Energy Convers. Manag. 74, 524–534. doi:10.1016/j.enconman.2013.07.024

Ou, L., Brown, T.R., Thilakaratne, R., Hu, G., Brown, R.C., 2014. Techno-economic analysis of co-located corn grain and corn stover ethanol plants. Biofuels Bioprod. Biorefining.

Patni, N., Pillai, S.G., Dwivedi, A.H., 2013. Wheat as a Promising Substitute of Corn for Bioethanol Production. Procedia Eng. 51, 355–362. doi:10.1016/j.proeng.2013.01.049

Paulová, L., Patáková, P., Rychtera, M., Melzoch, K., 2014. High solid fed-batch SSF with delayed inoculation for improved production of bioethanol from wheat straw. Fuel 122, 294–300. doi:10.1016/j.fuel.2014.01.020

Pejin, D., Mojović, L.J., Vučurović, V., Pejin, J., Denčić, S., Rakin, M., 2009. Fermentation of wheat and triticale hydrolysates: A comparative study. Fuel 88, 1625–1628. doi:10.1016/j.fuel.2009.01.011

Pejin, J.D., Mojović, L.V., Pejin, D.J., Kocić-Tanackov, S.D., Savić, D.S., Nikolić, S.B., Djukić-Vuković, A.P., 2015. Bioethanol production from triticale by simultaneous saccharification and fermentation with magnesium or calcium ions addition. Fuel 142, 58–64. doi:10.1016/j.fuel.2014.10.077

Petersen, A.M., 2015. Integration of Second Generation Biofuel Production into Existing Industrial Processes for Short Term Commercial Implementation. Stellenbosch: University of Stellenbosch, Stellenbosch.

Petersen, A.M., Aneke, M.C., Görgens, J.F., others, 2014. Techno-economic comparison of ethanol and electricity coproduction schemes from sugarcane residues at existing sugar mills in Southern Africa. Biotechnol Biofuels 7, 105.

Pieters, R.J., 2016. Potable ethanol production from raw corn using simultaneous saccharification and fermentation. Stellenbosch: University of Stellenbosch, Stellenbosch.

Pineo, C., Basson, L., Janse van Vuuren, P., Millson, C., Petrie, J., 2015. Regional Resource Flow Model: Grain Sector Report. GreenCape Sector Development Agency, Cape Town.

POET, 2015. POET's Technology Advantage - POET [WWW Document]. URL http://www.poet.com/advantage (accessed 11.16.15).

Ports.com, 2014. Sea route & distance - ports.com [WWW Document]. URL http://ports.com/sea-route/#/?a=16689&b=3848&c=Richards%20Bay,%20South%20Africa&d=Port%20of%20Saldanha,%20South%20Africa (accessed 2.22.15).

Pradhan, A., Mbohwa, C., 2014. Development of biofuels in South Africa: Challenges and opportunities. Renew. Sustain. Energy Rev. 39, 1089–1100. doi:10.1016/j.rser.2014.07.131

Richardson, J.W., Lemmer, W.J., Outlaw, J.L., 2007. Bio-ethanol production from wheat in the winter rainfall region of South Africa: a quantitative risk analysis. Int. Food Agribus. Manag. Rev. 10, 181–204.

Richard, T., Chisti, Y., Somerville, C.R., Blanch, H., Babcock, B., Gheewala, S.H., Zilberman, D., 2012. Ask the Experts: The food versus fuel debate. Biofuels 3, 635–648. doi:10.4155/bfs.12.59

SAFEX-JSE, 2015. JSE | Download Files [WWW Document]. URL http://www.jse.co.za/DownloadFiles.aspx?RequestedNode=DownloadableDocuments/Safex/agriculture.stats/ (accessed 4.11.14).

Sánchez, Ó.J., Cardona, C.A., 2008. Trends in biotechnological production of fuel ethanol from different feedstocks. Bioresour. Technol. 99, 5270–5295. doi:10.1016/j.biortech.2007.11.013

SARS, 2015. Corporate Income Tax [WWW Document]. URL http://www.sars.gov.za/TaxTypes/CIT/Pages/default.aspx (accessed 11.18.15).

SASA, 2015. SASA | Facts and Figures [WWW Document]. URL http://www.sasa.org.za/sugar_industry/FactsandFigures.aspx (accessed 11.4.15).

Sharma, V., Rausch, K.D., Graeber, J.V., Schmidt, S.J., Buriak, P., Tumbleson, M.E., Singh, V., 2009. Effect of Resistant Starch on Hydrolysis and Fermentation of Corn Starch for Ethanol. Appl. Biochem. Biotechnol. 160, 800–811. doi:10.1007/s12010-009-8651-7

Shewale, S.D., Pandit, A.B., 2009. Enzymatic production of glucose from different qualities of grain sorghum and application of ultrasound to enhance the yield. Carbohydr. Res. 344, 52–60. doi:10.1016/j.carres.2008.10.006

Singh, V., Moreau, R.A., Doner, L.W., Eckhoff, S.R., Hicks, K.B., 1999. Recovery of fiber in the corn dry-grind ethanol process: A feedstock for valuable coproducts. Cereal Chem. 76, 868–872.

Smith, A.M., Denyer, K., Martin, C., 1997. The synthesis of the starch granule. Annu. Rev. Plant Biol. 48, 67–87.

South African Reserve Bank, 2015. Selected historical rates - South African Reserve Bank [WWW Document]. URL https://www.resbank.co.za/Research/Rates/Pages/SelectedHistoricalExchangeAndInterestRates.aspx (accessed 11.18.15).

Statistics South Africa, 2015. Economy | Statistics South Africa [WWW Document]. URL http://www.statssa.gov.za/?page_id=735&id=3 (accessed 11.18.15).

Tait, B., 2014. The biofuels Pricing and Incentive Models.

Tait, B., 2011. Biofuels Pricing and Manufacturing Economics Study.
Ten Cate, S., 2015a. DDGS Prys. [Personal Communication]
Ten Cate, S., 2015b. Korog prys. [Personal Communication]
Tsupko, Y.V., 2009. Investigation into the suitability of spring triticale ($\times$ Triticosecale Wittmack) for bio-ethanol production in the Western Cape. Stellenbosch: University of Stellenbosch.
Turton, R., 2009. Analysis, synthesis, and design of chemical processes, 3rd ed.. ed. Prentice Hall, Upper Saddle River, NJ.
UCSB Science Line, 2015. How does ethyl alcohol kill bacteria? [WWW Document]. URL http://scienceline.ucsb.edu/getkey.php?key=2160 (accessed 11.16.15).
U.S. Energy Information Administration (EIA), 2014. How much carbon dioxide is produced by burning gasoline and diesel fuel? - FAQ - U.S. Energy Information Administration (EIA) [WWW Document]. URL http://www.eia.gov/tools/faqs/faq.cfm?id=307&t=11 (accessed 1.23.15).
U.S. Grain council, 2012. A guide to Distiller's Dried Grains with Solubles (DDGS).
Van Rensburg, E., 2014. Inoculum Yeast. [Personal Communication]
Van Rensburg, E., den Haan, R., la Grange, D.C., Volschenk, H., van Zyl, W.H., Görgens, J.F., 2013. Engineering Recombinant Organisms for Next Generation Ethanol Production. p. 91.
Verband der Automobilindustrie, 2012. Commercial Vehicles – Driving The Future. Berlin.
Vohra, M., Manwar, J., Manmode, R., Padgilwar, S., Patil, S., 2013. Bioethanol production: Feedstock and current technologies. J. Environ. Chem. Eng. doi:10.1016/j.jece.2013.10.013
Vučurović, V.M., Pejin, D.J., 2007. Production of bioethanol from triticale. Zb. Matice Srp. Za Prir. Nauke 285–291.
Walas, S.M., 1990. Chemical Process Equipment - Selection and Design, 1st ed. Butterworth-Heinemann, Boston.
Wang, D., Bean, S., McLaren, J., Seib, P., Madl, R., Tuinstra, M., Shi, Y., Lenz, M., Wu, X., Zhao, R., 2008. Grain sorghum is a viable feedstock for ethanol production. J. Ind. Microbiol. Biotechnol. 35, 313–320. doi:10.1007/s10295-008-0313-1
Wang, M., Wu, M., Huo, H., 2007. Life-cycle energy and greenhouse gas emission impacts of different corn ethanol plant types. Environ. Res. Lett. 2, 024001. doi:10.1088/1748-9326/2/2/024001
Wang, P., Singh, V., Xue, H., Johnston, D.B., Rausch, K.D., Tumbleson, M.E., 2007. Comparison of raw starch hydrolyzing enzyme with conventional liquefaction and saccharification enzymes in dry-grind corn processing. Cereal Chem. 84, 10–14.
Wang, S., Thomas, K.C., Ingledew, W.M., Sosulski, K., Sosulski, F.W., 1998. Production of fuel ethanol from rye and triticale by very-high-gravity (VHG) fermentation. Appl. Biochem. Biotechnol. 69, 157–175.
Wang, S., Thomas, K.C., Ingledew, W.M., Sosulski, K., Sosulski, F.W., 1997. Rye and triticale as feedstock for fuel ethanol production. Cereal Chem. 74, 621–625.
Wang, S., Thomas, K.C., Sosulski, K., Ingledew, W.M., Sosulski, F.W., 1999. Grain pearling and very high gravity (VHG) fermentation technologies for fuel alcohol production from rye and triticale. Process Biochem. 34, 421–428.
Water Rhapsody, 2014. Cape Town Water and Sewerage tariffs 2015Water Rhapsody | Water Rhapsody.
Willem Botes, 2014. Triticale for bioethanol production. [Personal Communication]

Zhao, R., Bean, S.R., Wang, D., Park, S.H., Schober, T.J., Wilson, J.D., 2009. Small-scale mashing procedure for predicting ethanol yield of sorghum grain. J. Cereal Sci. 49, 230–238. doi:10.1016/j.jcs.2008.10.006

8. APPENDIX A: ASPEN MODELS

In this section the Aspen process flow diagrams as well as the stream tables are presented for all 4 models.

Model 1

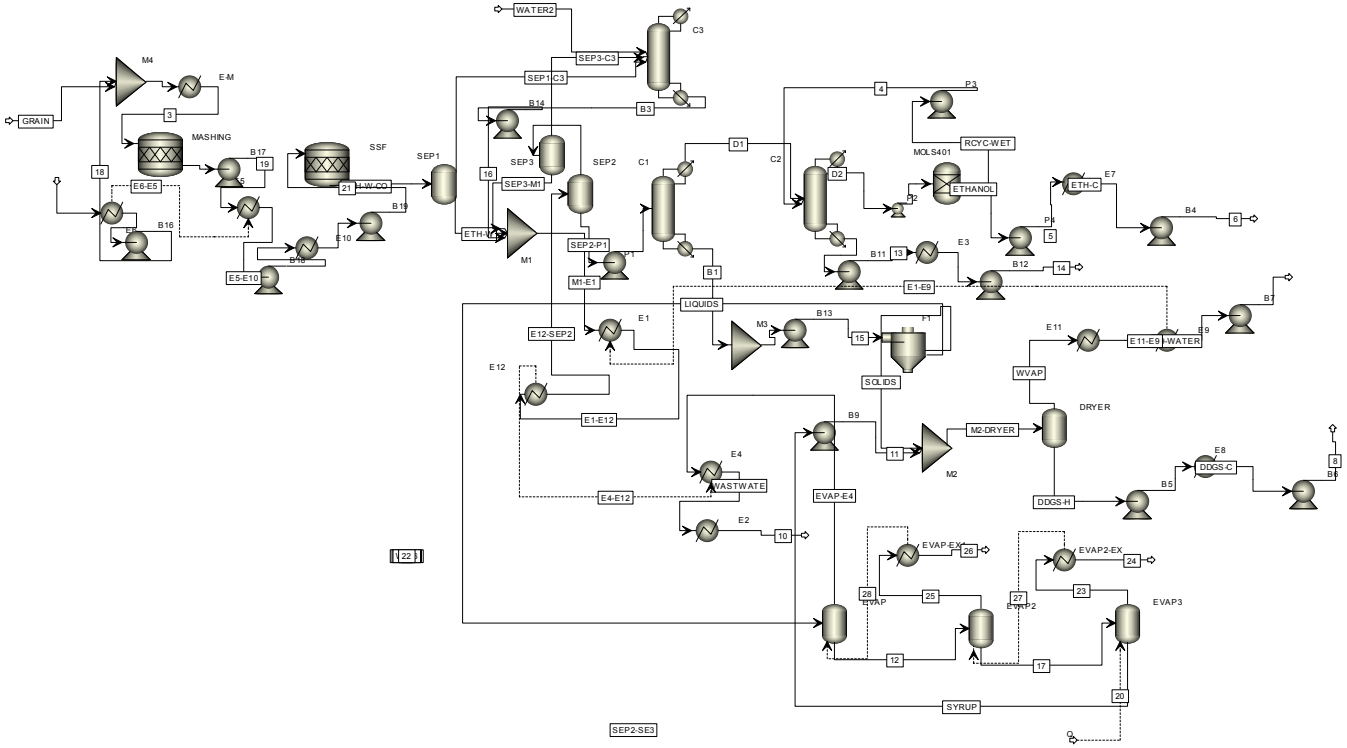

FIGURE 38: MODEL 1 ASPEN FLOW DIAGRAM

TABLE 9: ASPEN MODEL 1 LIQUID STREAM TABLE 1

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18
Temperature C	70	48.5	60	69.8	70.2	35	102.4	35	45	35	83	100.1	100	35	101.5	42.5	93.6	50
Pressure bar	1.5	1.013	1.013	1.5	1.5	1.5	1.5	1.5	1.5	1.013	1.5	1.013	1.5	1.5	1.5	1.5	0.8	1.5
Vapor Frac	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mole Flow kmol/hr	538.527	8867.417	8867.417	193.908	344.619	344.619	113.475	113.475	1253.697	2087.265	121.158	4269.933	1632.937	1632.937	7603.212	570.89	2190.178	8587.877
Mass Flow kg/hr	21766.47	159749	159749	5905.656	15860.82	15860.82	3512.36	3512.36	22585.75	37602.85	3410.175	78151.55	29419.03	29419.03	138442.3	10594.11	40684.15	154713
Volume Flow cum/hr	28.642	164.527	166.488	7.348	21.318	20.155	2.729	2.526	23.178	38.204	2.755	84.308	32.038	29.891	150.057	10.966	43.178	159.582
Enthalpy Gcal/hr	-35.213	-601.568	-599.708	-12.864	-22.361	-22.746	-9.653	-9.858	-85.131	-142.11	-9.887	-287.264	-109.196	-111.176	-510.29	-38.782	-148.471	-582.372
Mass Flow kg/hr																		
WATER	1953.928	159749	159749	1944.158	9.77	9.77	1881.175	1881.175	22585.68	37602.54	2046.309	76787.63	29417.05	29417.05	136810.7	10085.77	39320.28	154713
ETHANOL	19807.49	0	0	3961.498	15845.99	15845.99	0.001	0.001	0.071	0.312	0	0.051	1.981	1.981	0.434	497.691	0.004	0
CARBO-01	5.055	0	0	0	5.055	5.055	0	0	0	0	0	0	0	0	0	10.654	0	0
GLUCOSE	0	0	0	0	0	0	1631.184	1631.184	0	0	1363.866	1363.866	0	0	1631.185	0	1363.866	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

TABLE 10: ASPEN MODEL 1 SOLID STREAM TABLE 1

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18
Mole Flow kmol/hr	0	441.689	441.689	0	0	0	260.604	260.604	0	0	0	0	0	0	260.604	0	0	0
Mass Flow kg/hr	0	41578	41578	0	0	0	12216.58	12216.58	0	0	0	0	0	0	12216.58	0	0	0
Volume Flow cum/hr	0	11.932	11.932	0	0	0	7.04	7.04	0	0	0	0	0	0	7.04	0	0	0
Enthalpy Gcal/hr		-49.827	-49.772				-7.491	-7.683							-7.494			
Mass Flow kg/hr																		
WATER	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHANOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CARBO-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	0	29658	29658	0	0	0	296.58	296.58	0	0	0	0	0	0	296.58	0	0	0
CELLU-01	0	3891	3891	0	0	0	3891	3891	0	0	0	0	0	0	3891	0	0	0
XYLAN	0	3104	3104	0	0	0	3104	3104	0	0	0	0	0	0	3104	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	0	4925	4925	0	0	0	4925	4925	0	0	0	0	0	0	4925	0	0	0
CELLULOS																		

TABLE 11: ASPEN MODEL 1 LIQUID STREAM TABLE 2

	19	21	22	23	24	25	26	B1	B2	B3	C1-ETH	D1	D2	D3	DDGS-C	DDGS-H	E1-E12	E2-F1
Temperature C																		
Pressure bar	60	30	38.9	83	100	93.6	100	101.5	100	42.4	86	95.2	70	28.6	35	102.4	38.3	101.5
Vapor Frac	1.5	1.5	1.5	0.5	1.013	0.8	1.013	1.076	1.013	1.013	1.5	1.013	1.013	1.013	1.013	1.013	1.013	1.076
Mole Flow kmol/hr	0	0	0	1	0	1	0	0	0	0	0	1	0	1	0	0	0	0
Mass Flow kg/hr	8867.417	8867.417	8867.417	2069.02	2069.02	2079.755	2079.755	7603.212	1632.937	570.89	9580.773	1977.561	538.527	357.818	113.475	113.475	9706.426	7603.212
Volume Flow cum/hr	189110.4	189110.4	189110.4	37273.97	37273.97	37467.4	37467.4	138442.3	29419.03	10594.11	183722.3	45279.92	21766.47	15388.16	3512.36	3512.36	187062.7	138442.3
Enthalpy Gcal/hr	177.579	172.061	173.636	122519.5	40.591	79275.43	40.802	150.055	32.038	10.965	200.578	59779.78	28.64	8859.824	2.526	2.729	223.931	150.055
Mass Flow kg/hr	-641.891	-647.246	-645.681	-118.535	-138.358	-118.97	-139.076	-510.292	-109.197	-38.783	-644.581	-112.324	-35.213	-33.117	-9.858	-9.653	-661.999	-510.292
WATER																		
ETHANOL	156486.7	156486.7	156486.7	37273.97	37273.97	37467.35	37467.35	136810.7	29417.05	10085.77	166237.6	29426.89	1953.928	249.111	1881.175	1881.175	167801.7	136810.7
CARBO-01	0	0	0	0.004	0.004	0.047	0.047	0.434	1.981	497.691	15848.41	15847.97	19807.49	2.101	0.001	0.001	17353.33	0.434
GLUCOSE	0	0	0	0	0	0	0	0	0	10.654	5.055	5.055	5.055	15136.95	0	0	276.465	0
AIR	32623.7	32623.7	32623.7	0	0	0	0	1631.185	0	0	1631.185	0	0	0	1631.184	1631.184	1631.185	1631.185
STARCH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULOS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

TABLE 12: ASPEN MODEL 1 SOLID STREAM TABLE 1

	19	21	22	23	24	25	26	B1	B2	B3	C1-ETH	D1	D2	D3	DDGS-C	DDGS-H	E1-E12	E2-F1
Mole Flow kmol/hr	260.604	260.604	260.604	0	0	0	0	260.604	0	0	260.604	0	0	0	260.604	260.604	260.604	260.604
Mass Flow kg/hr	12216.58	12216.58	12216.58	0	0	0	0	12216.58	0	0	12216.58	0	0	0	12216.58	12216.58	12216.58	12216.58
Volume Flow cum/hr	7.04	7.04	7.04	0	0	0	0	7.04	0	0	7.04	0	0	0	7.04	7.04	7.04	7.04
Enthalpy Gcal/hr	-7.612	-7.697	-7.672					-7.494			-7.538				-7.683	-7.491	-7.674	-7.494
Mass Flow kg/hr																		
WATER	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHANOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CARBO-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	296.58	296.58	296.58	0	0	0	0	296.58	0	0	296.58	0	0	0	296.58	296.58	296.58	296.58
CELLU-01	3891	3891	3891	0	0	0	0	3891	0	0	3891	0	0	0	3891	3891	3891	3891
XYLAN	3104	3104	3104	0	0	0	0	3104	0	0	3104	0	0	0	3104	3104	3104	3104
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	4925	4925	4925	0	0	0	0	4925	0	0	4925	0	0	0	4925	4925	4925	4925
CELLULOS																		

TABLE 13: ASPEN MODEL 1 LIQUID STREAM TABLE 3

	E3-WAST	E5-E10	E9-WATER	E10-SSF	E11-E9	E12-SEP2	ETH-C	ETH-W	ETH-W-CC	ETHANOL	EVAP-E4	GRAIN	LIQUE	LIQUIDS	M1-E1	M2-DRYER	RCYC-WET	SEP1-C3
Temperature C																		
Pressure bar	35	38.9	45	30	100	86	35	30	30	70.1	100.1	25	60	100.1	31.3	99.5	70.1	30
Vapor Frac	1	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013
Mole Flow kmol/hr	0	0	0	0	0	0.003	0	0	0.038	0	1	0	0	0.003	0	0	0	1
Mass Flow kg/hr	1632.937	8867.417	1253.697	8867.417	1253.697	9706.426	344.619	9025.552	9383.507	344.619	2087.265	279.54	8867.417	6357.198	9706.426	1367.173	193.908	357.955
Volume Flow cum/hr	29419.03	189110.4	22585.75	189110.4	22585.75	187062.7	15860.82	173721.9	189110.4	15860.82	37602.85	5036	189110.4	115754.4	187062.7	26098.11	5905.656	15388.54
Enthalpy Gcal/hr	29.89	173.633	23.178	172.058	24.595	1040.559	20.153	178.346	9082.573	21.315	63921.98	5.067	177.576	698.334	192.64	27.431	7.313	8904.227
Mass Flow kg/hr	-111.177	-645.684	-85.131	-647.249	-83.837	-652.885	-22.746	-617.132	-650.03	-22.362	-119.29	-19.082	-641.894	-426.66	-663.273	-93.511	-12.864	-32.898
WATER																		
ETHANOL	29417.05	156486.7	22585.68	156486.7	22585.68	167801.7	9.77	156226	156486.7	9.77	37602.54	5036	156486.7	114390.2	167801.7	24466.85	1944.158	260.673
CARBO-01	1.981	0	0.071	0	0.071	17353.33	15845.99	15600.97	15850.51	15845.99	0.312	0	0	0.363	17353.33	0.071	3961.498	249.54
GLUCOSE	0	0	0	0	0	276.465	5.055	263.681	15142.01	5.055	0	0	0	0	276.465	0	0	14878.33
AIR	0	32623.7	0	32623.7	0	1631.185	0	1631.185	1631.185	0	0	0	32623.7	1363.866	1631.185	1631.185	0	0
STARCH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULOS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

TABLE 14: ASPEN MODEL 1 SOLID STREAM TABLE 3

	E3-WAST	E5-E10	E9-WATER	E10-SSF	E11-E9	E12-SEP2	ETH-C	ETH-W	ETH-W-CC	ETHANOL	EVAP-E4	GRAIN	LIQUE	LIQUIDS	M1-E1	M2-DRYER	RCYC-WET	SEP1-C3
Mole Flow kmol/hr	0	260.604	0	260.604	0	260.604	0	260.604	260.604	0	0	441.689	260.604	0	260.604	260.604	0	0
Mass Flow kg/hr	0	12216.58	0	12216.58	0	12216.58	0	12216.58	12216.58	0	0	41578	12216.58	0	12216.58	12216.58	0	0
Volume Flow cum/hr	0	7.04	0	7.04	0	7.04	0	7.04	7.04	0	0	11.932	7.04	0	7.04	7.04	0	0
Enthalpy Gcal/hr		-7.672		-7.697		-7.538		-7.697	-7.697			-49.94	-7.612		-7.694	-7.5		
Mass Flow kg/hr																		
WATER	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHANOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CARBO-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	0	296.58	0	296.58	0	296.58	0	296.58	296.58	0	0	29658	296.58	0	296.58	296.58	0	0
CELLU-01	0	3891	0	3891	0	3891	0	3891	3891	0	0	3891	3891	0	3891	3891	0	0
XYLAN	0	3104	0	3104	0	3104	0	3104	3104	0	0	3104	3104	0	3104	3104	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	0	4925	0	4925	0	4925	0	4925	4925	0	0	4925	4925	0	4925	4925	0	0
CELLULOS																		

TABLE 15: ASPEN MODEL 1 LIQUID STREAM TABLE 4

	SEP2-P1	SEP2-SE3	SEP3-C3	SEP3-M1	SOLIDS	SYRUP	W-E6	WASTWA	WATER	WATER2	WVAP
Temperature C											
Pressure bar	86	86	70	70	100.1	83	50	100	25	25	102.4
Vapor Frac	0.841	0.841	1.013	1.013	1.013	0.5	1.013	1.013	1.013	1.013	1.013
Mole Flow kmol/hr	0	1	1	0	0.003	0	0	0.544	0	0	1
Mass Flow kg/hr	9580.773	125.653	15.67	109.983	1246.015	121.158	8587.877	2087.265	8587.877	555.084	1253.697
Volume Flow cum/hr	183722.3	3340.41	593.735	2746.675	22687.93	3410.175	154713	37602.85	154713	10000	22585.75
Enthalpy Gcal/hr	200.573	4461.507	441.219	3.247	136.874	2.755	159.58	34793.16	155.654	10.061	38630.68
Mass Flow kg/hr	-644.585	-7.348	-1.11	-7.355	-83.626	-9.888	-582.374	-128.54	-586.227	-37.891	-71.627
WATER											
ETHANOL	166237.6	1564.074	74.204	1489.871	22420.55	2046.309	154713	37602.54	154713	10000	22585.68
CARBO-01	15848.41	1504.925	250.251	1254.673	0.071	0	0	0.312	0	0	0.071
GLUCOSE	5.055	271.411	269.28	2.131	0	0	0	0	0	0	0
AIR	1631.185	0	0	0	267.319	1363.866	0	0	0	0	0
STARCH	0	0	0	0	0	0	0	0	0	0	0
CELLU-01	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	0	0	0	0	0	0	0	0	0	0	0
CELLULOS	0	0	0	0	0	0	0	0	0	0	0

TABLE 16: ASPEN MODEL 1 SOLID STREAM TABLE 4

	SEP2-P1	SEP2-SE3	SEP3-C3	SEP3-M1	SOLIDS	SYRUP	W-E6	WASTWA	WATER	WATER2	WVAP
Mole Flow kmol/hr	260.604	0	0	0	260.604	0	0	0	0	0	0
Mass Flow kg/hr	12216.58	0	0	0	12216.58	0	0	0	0	0	0
Volume Flow cum/hr	7.04	0	0	0	7.04	0	0	0	0	0	0
Enthalpy Gcal/hr	-7.538				-7.498						
Mass Flow kg/hr											
WATER	0	0	0	0	0	0	0	0	0	0	0
ETHANOL	0	0	0	0	0	0	0	0	0	0	0
CARBO-01	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0
STARCH	296.58	0	0	0	296.58	0	0	0	0	0	0
CELLU-01	3891	0	0	0	3891	0	0	0	0	0	0
XYLAN	3104	0	0	0	3104	0	0	0	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	4925	0	0	0	4925	0	0	0	0	0	0

Model 2

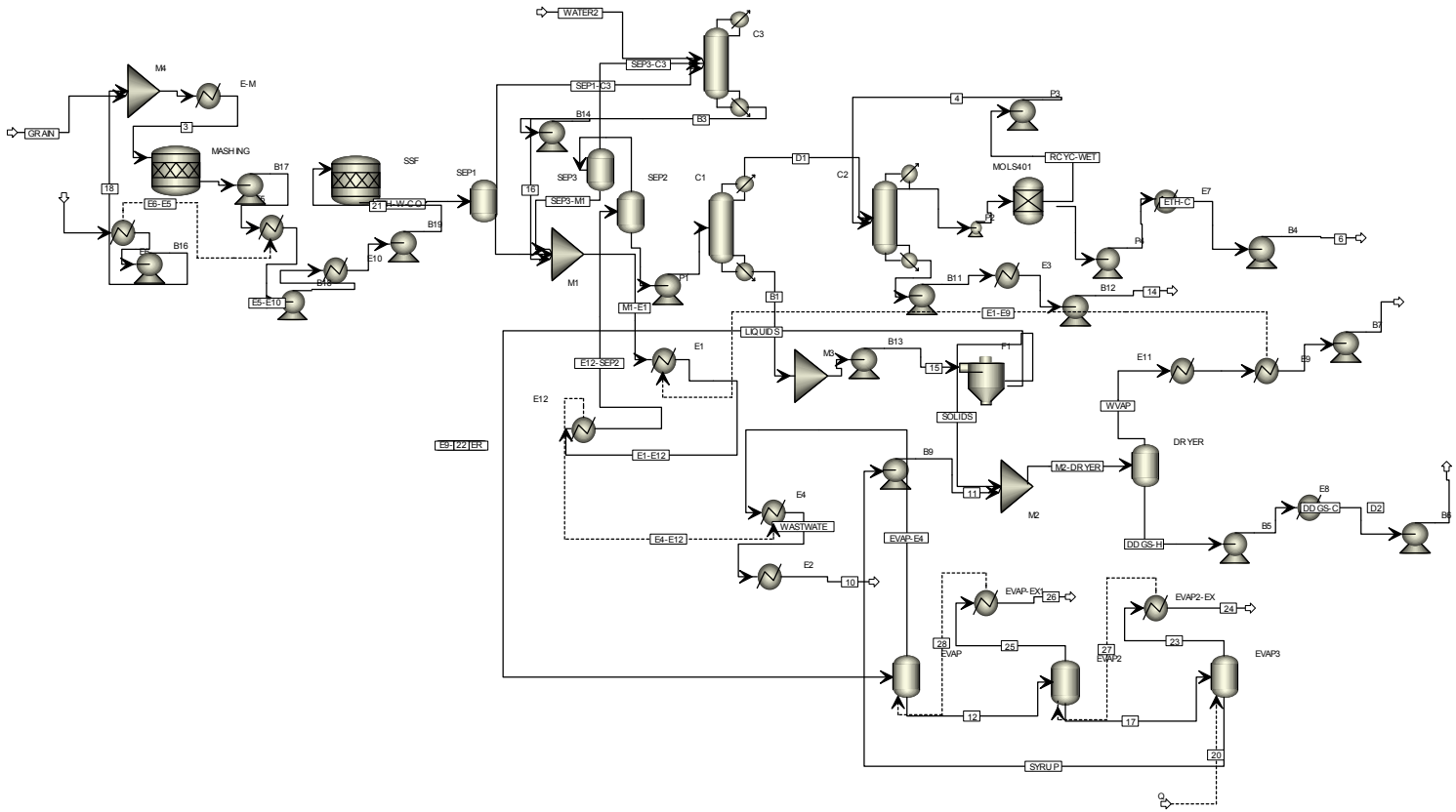

FIGURE 39: MODEL 2 ASPEN FLOW DIAGRAM

TABLE 17: ASPEN MODEL 2 LIQUID STREAM TABLE 1

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
Temperature C	70	48.5	60	69.8	70.2	35	102.1	35	45	35	82.5	100.1	100	35	101.5	42.2	93.6	50	57
Pressure bar	1.5	1.013	1.013	1.5	1.5	1.5	1.5	1.5	1.5	1.013	1.5	1.013	1.5	1.5	1.5	1.5	0.8	1.5	1.5
Vapor Frac	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mole Flow kmol/hr	522.192	8867.417	8867.417	187.996	334.196	334.196	124.637	124.637	1377.01	2060.738	164.356	4260.514	1583.436	1583.436	7658.543	570.556	2207.266	8587.877	8867.417
Mass Flow kg/hr	21107.15	159749	159749	5726.088	15381.06	15381.06	3668.949	3668.949	24807.29	37125.01	4135.927	77929.4	28527.21	28527.21	139394.7	10575.67	40939.52	154713	188220.7
Volume Flow cum/hr	27.774	164.527	166.488	7.124	20.673	19.545	2.946	2.728	25.458	37.719	3.589	84.101	31.067	28.985	151.124	10.938	43.488	159.582	176.689
Enthalpy Gcal/hr	-34.145	-601.568	-599.708	-12.472	-21.685	-22.058	-10.337	-10.553	-93.504	-140.304	-12.718	-286.56	-105.886	-107.806	-513.926	-38.763	-149.541	-582.372	-641.155
Mass Flow kg/hr																			
WATER	1894.056	159749	159749	1884.586	9.47	9.47	2087.194	2087.194	24807.19	37124.62	2830.369	76623.78	28525.29	28525.29	137812.4	10087.74	39633.95	154713	156585.6
ETHANOL	19207.51	0	0	3841.501	15366.01	15366.01	0.001	0.001	0.096	0.393	0	0.064	1.921	1.921	0.554	477.258	0.005	0	0
CARBO-01	5.587	0	0	0	5.587	5.587	0	0	0	0	0	0	0	0	0	10.672	0	0	0
GLUCOSE	0	0	0	0	0	0	1581.755	1581.755	0	0	1305.558	1305.558	0	0	1581.755	0	1305.558	0	31635.1
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULOS																			

TABLE 18: ASPEN MODEL 2 SOLID STREAM TABLE 1

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
Mole Flow kmol/hr	0	441.689	441.689	0	0	0	266.092	266.092	0	0	0	0	0	0	266.092	0	0	0	266.092
Mass Flow kg/hr	0	41578	41578	0	0	0	13106.32	13106.32	0	0	0	0	0	0	13106.32	0	0	0	13106.32
Volume Flow cum/hr	0	11.932	11.932	0	0	0	7.188	7.188	0	0	0	0	0	0	7.188	0	0	0	7.188
Enthalpy Gcal/hr		-49.827	-49.772				-8.767	-8.962							-8.769				-8.898
Mass Flow kg/hr																			
WATER	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHANOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CARBO-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	0	29658	29658	0	0	0	1186.32	1186.32	0	0	0	0	0	0	1186.32	0	0	0	1186.32
CELLU-01	0	3891	3891	0	0	0	3891	3891	0	0	0	0	0	0	3891	0	0	0	3891
XYLAN	0	3104	3104	0	0	0	3104	3104	0	0	0	0	0	0	3104	0	0	0	3104
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	0	4925	4925	0	0	0	4925	4925	0	0	0	0	0	0	4925	0	0	0	4925
CELLULOS																			

TABLE 19: ASPEN MODEL 2 LIQUID STREAM TABLE 2

	21	22	23	24	25	26	B1	B2	B3	C1-ETH	D1	D2	D3	DDGS-C	DDGS-H	E1-E12	E2-F1	E3-WAST	E5-E10
Temperature C	30	35.8	82.5	100	93.6	100	101.5	100	42.2	86	95.2	70	28.4	35	102.1	39	101.5	35	35.7
Pressure bar	1.5	1.5	0.5	1.013	0.8	1.013	1.076	1.013	1.013	1.5	1.013	1.013	1.013	1.013	1.013	1.013	1.076	1	1.013
Vapor Frac	0	0	1	0	1	0	0	0	0	0	1	0	1	0	0	0	0	0	0
Mole Flow kmol/hr	8867.417	8867.417	2042.91	2042.91	2053.249	2053.249	7658.543	1583.436	570.556	9576.18	1917.636	522.192	346.772	124.637	124.637	9687.71	7658.543	1583.436	8867.417
Mass Flow kg/hr	188220.7	188220.7	36803.59	36803.59	36989.89	36989.89	139394.7	28527.21	10575.67	183303.1	43908.36	21107.15	14917.59	3668.949	3668.949	186264.8	139394.7	28527.21	188220.7
Volume Flow cum/hr	171.761	172.772	120807.3	40.079	78264.06	40.282	151.122	31.066	10.938	200.018	57968.16	27.773	8579.94	2.728	2.945	225.749	151.122	28.984	172.769
Enthalpy Gcal/hr	-645.951	-644.943	-117.047	-136.612	-117.454	-137.303	-513.928	-105.887	-38.764	-644.234	-108.921	-34.146	-32.101	-10.553	-10.337	-660.571	-513.928	-107.807	-644.946
Mass Flow kg/hr																			
WATER	156585.6	156585.6	36803.59	36803.59	36989.83	36989.83	137812.4	28525.29	10087.74	166347.3	28534.85	1894.056	238.31	2087.194	2087.194	167736.8	137812.4	28525.29	156585.6
ETHANOL	0	0	0.005	0.005	0.059	0.059	0.554	1.921	477.258	15368.48	15367.93	19207.51	1.709	0.001	0.001	16672.02	0.554	1.921	0
CARBO-01	0	0	0	0	0	0	0	0	10.672	5.587	5.587	5.587	14677.57	0	0	274.16	0	0	0
GLUCOSE	31635.1	31635.1	0	0	0	0	1581.755	0	0	1581.755	0	0	0	1581.755	1581.755	1581.755	1581.755	0	31635.1
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULOS																			

TABLE 20: ASPEN MODEL 2 SOLID STREAM TABLE 2

	21	22	23	24	25	26	B1	B2	B3	C1-ETH	D1	D2	D3	DDGS-C	DDGS-H	E1-E12	E2-F1	E3-WAST	E5-E10
Mole Flow kmol/hr	266.092	266.092	0	0	0	0	266.092	0	0	266.092	0	0	0	266.092	266.092	266.092	266.092	0	266.092
Mass Flow kg/hr	13106.32	13106.32	0	0	0	0	13106.32	0	0	13106.32	0	0	0	13106.32	13106.32	13106.32	13106.32	0	13106.32
Volume Flow cum/hr	7.188	7.188	0	0	0	0	7.188	0	0	7.188	0	0	0	7.188	7.188	7.188	7.188	0	7.188
Enthalpy Gcal/hr	-8.977	-8.96					-8.769			-8.814				-8.962	-8.767	-8.951	-8.769		-8.96
Mass Flow kg/hr																			
WATER	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHANOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CARBO-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	1186.32	1186.32	0	0	0	0	1186.32	0	0	1186.32	0	0	0	1186.32	1186.32	1186.32	1186.32	0	1186.32
CELLU-01	3891	3891	0	0	0	0	3891	0	0	3891	0	0	0	3891	3891	3891	3891	0	3891
XYLAN	3104	3104	0	0	0	0	3104	0	0	3104	0	0	0	3104	3104	3104	3104	0	3104
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	4925	4925	0	0	0	0	4925	0	0	4925	0	0	0	4925	4925	4925	4925	0	4925
CELLULOS																			

TABLE 21: ASPEN MODEL 2 LIQUID STREAM TABLE 3

	E9-WATER	E10-SSF	E11-E9	E12-SEP2	ETH-C	ETH-W	ETH-W-CC	ETHANOL	EVAP-E4	GRAIN	LIQUE	LIQUIDS	M1-E1	M2-DRYER	RCYC-WET	SEP1-C3	SEP2-P1	SEP2-SE3
Temperature C	45	30	100	86	35	30	30	70.1	100.1	25	57	100	31.2	99.1	70.1	30	86	86
Pressure bar	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	0.841	0.841
Vapor Frac	0	0	0	0.003	0	0	0.037	0	1	0	0	0.003	0	0	0	1	0	1
Mole Flow kmol/hr	1377.01	8867.417	1377.01	9687.71	334.196	9021.026	9367.868	334.196	2060.738	279.54	8867.417	6321.252	9687.71	1501.647	187.996	346.841	9576.18	111.53
Mass Flow kg/hr	24807.29	188220.7	24807.29	186264.8	15381.06	173310.5	188220.7	15381.06	37125.01	5036	188220.7	115054.4	186264.8	28476.24	5726.088	14910.15	183303.1	2961.669
Volume Flow cum/hr	25.458	171.758	27.015	1006.235	19.544	177.848	8805.627	20.67	63109.26	5.067	176.686	692.804	191.766	30.038	7.091	8627.779	200.013	3960.047
Enthalpy Gcal/hr	-93.505	-645.954	-92.083	-651.609	-22.058	-616.773	-648.654	-21.686	-117.774	-19.082	-641.158	-424.184	-661.97	-102.453	-12.472	-31.881	-644.238	-6.546
Mass Flow kg/hr																		
WATER	24807.19	156585.6	24807.19	167736.8	9.47	156332.7	156585.6	9.47	37124.62	5036	156585.6	113748.4	167736.8	26894.38	1884.586	252.829	166347.3	1389.557
ETHANOL	0.096	0	0.096	16672.02	15366.01	15134.36	15370.19	15366.01	0.393	0	0	0.457	16672.02	0.097	3841.501	235.832	15368.48	1303.539
CARBO-01	0	0	0	274.16	5.587	261.672	14683.16	5.587	0	0	0	0	274.16	0	0	14421.49	5.587	268.573
GLUCOSE	0	31635.1	0	1581.755	0	1581.755	1581.755	0	0	0	31635.1	1305.558	1581.755	1581.755	0	0	1581.755	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULOS																		

TABLE 22: ASPEN MODEL 2 SOLID STREAM TABLE 3

	E9-WATER	E10-SSF	E11-E9	E12-SEP2	ETH-C	ETH-W	ETH-W-CC	ETHANOL	EVAP-E4	GRAIN	LIQUE	LIQUIDS	M1-E1	M2-DRYER	RCYC-WET	SEP1-C3	SEP2-P1	SEP2-SE3
Mole Flow kmol/hr	0	266.092	0	266.092	0	266.092	266.092	0	0	441.689	266.092	0	266.092	266.092	0	0	266.092	0
Mass Flow kg/hr	0	13106.32	0	13106.32	0	13106.32	13106.32	0	0	41578	13106.32	0	13106.32	13106.32	0	0	13106.32	0
Volume Flow cum/hr	0	7.188	0	7.188	0	7.188	7.188	0	0	11.932	7.188	0	7.188	7.188	0	0	7.188	0
Enthalpy Gcal/hr		-8.977		-8.814		-8.977	-8.977			-49.94	-8.898		-8.973	-8.776			-8.814	
Mass Flow kg/hr																		
WATER	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHANOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CARBO-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	0	1186.32	0	1186.32	0	1186.32	1186.32	0	0	29658	1186.32	0	1186.32	1186.32	0	0	1186.32	0
CELLU-01	0	3891	0	3891	0	3891	3891	0	0	3891	3891	0	3891	3891	0	0	3891	0
XYLAN	0	3104	0	3104	0	3104	3104	0	0	3104	3104	0	3104	3104	0	0	3104	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	0	4925	0	4925	0	4925	4925	0	0	4925	4925	0	4925	4925	0	0	4925	0
CELLULOS																		

TABLE 23: ASPEN MODEL 2 LIQUID STREAM TABLE 4

	SEP3-C3	SEP3-M1	SOLIDS	SYRUP	W-E6	WASTWA	WATER	WATER2	WVAP
Temperature C	70	70	100	82.5	50	100	25	25	102.1
Pressure bar	1.013	1.013	1.013	0.5	1.013	1.013	1.013	1.013	1.013
Vapor Frac	1	0	0.003	0	0	0.546	0	0	1
Mole Flow kmol/hr	15.403	96.127	1337.291	164.356	8587.877	2060.738	8587.877	555.084	1377.01
Mass Flow kg/hr	583.108	2378.561	24340.31	4135.927	154713	37125.01	154713	10000	24807.29
Volume Flow cum/hr	433.711	2.804	146.566	3.589	159.58	34456.52	155.654	10.061	42397.92
Enthalpy Gcal/hr	-1.093	-6.431	-89.738	-12.718	-582.374	-126.873	-586.227	-37.891	-78.676
Mass Flow kg/hr									
WATER	73.216	1316.341	24064.02	2830.369	154713	37124.62	154713	10000	24807.19
ETHANOL	243.135	1060.404	0.097	0	0	0.393	0	0	0.096
CARBO-01	266.757	1.816	0	0	0	0	0	0	0
GLUCOSE	0	0	276.197	1305.558	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0
STARCH	0	0	0	0	0	0	0	0	0
CELLU-01	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0
UREA	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0
PROTEIN	0	0	0	0	0	0	0	0	0
CELLULOS									

TABLE 24: ASPEN MODEL 2 SOLID STREAM TABLE 4

	SEP3-C3	SEP3-M1	SOLIDS	SYRUP	W-E6	WASTWA	WATER	WATER2	WVAP
Mole Flow kmol/hr	0	0	266.092	0	0	0	0	0	0
Mass Flow kg/hr	0	0	13106.32	0	0	0	0	0	0
Volume Flow cum/hr	0	0	7.188	0	0	0	0	0	0
Enthalpy Gcal/hr			-8.773						
Mass Flow kg/hr									
WATER	0	0	0	0	0	0	0	0	0
ETHANOL	0	0	0	0	0	0	0	0	0
CARBO-01	0	0	0	0	0	0	0	0	0
GLUCOSE	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0
STARCH	0	0	1186.32	0	0	0	0	0	0
CELLU-01	0	0	3891	0	0	0	0	0	0
XYLAN	0	0	3104	0	0	0	0	0	0
UREA	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0
PROTEIN	0	0	4925	0	0	0	0	0	0
CELLULOS									

Model 3

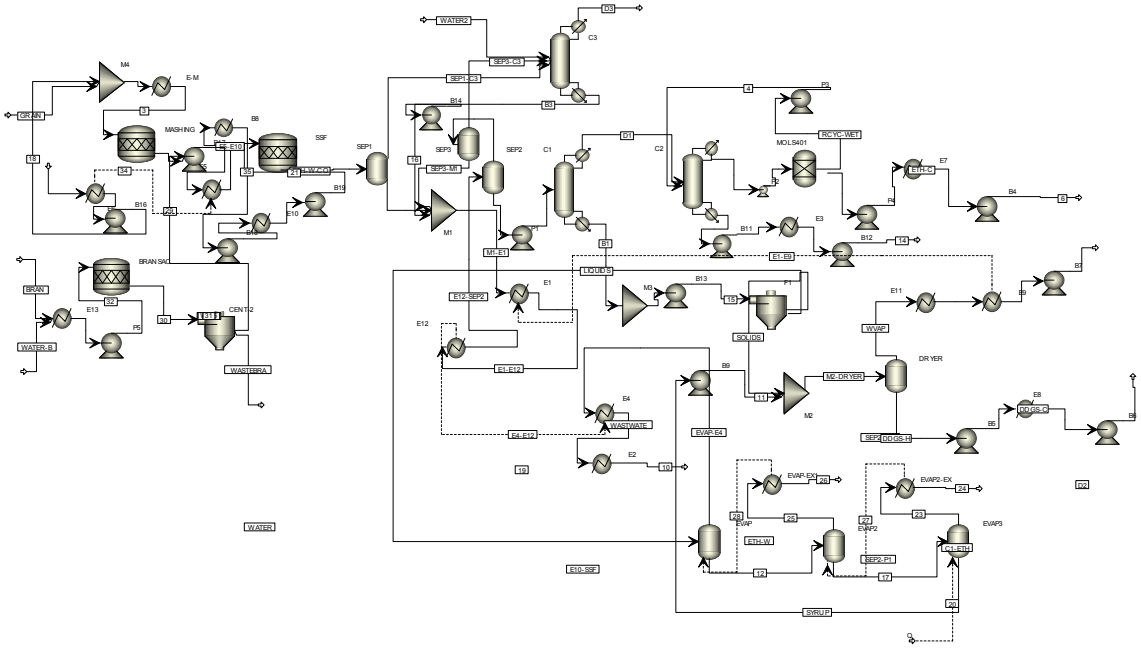

FIGURE 40: MODEL 3 ASPEN FLOW DIAGRAM

TABLE 25: ASPEN MODEL 3 LIQUID STREAM TABLE 1

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
Temperature C	70	48.7	60	69.8	70.2	35	104.7	35	45	35	82.7	100.1	100	35	101.5	42.4	93.6	50	57.3
Pressure bar	1.5	1.013	1.013	1.5	1.5	1.5	1.5	1.5	1.5	1.013	1.5	1.013	1.5	1.5	1.5	1.5	0.8	1.5	1.5
Vapor Frac	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mole Flow kmol/hr	533.159	7955.469	7955.469	191.975	341.185	341.185	78.535	78.535	655.426	2281.609	183.76	4717.552	1616.663	1616.663	7549.362	570.786	2444.815	7729.05	8795.388
Mass Flow kg/hr	21549.54	143320	143320	5846.793	15702.75	15702.75	2960.948	2960.948	11807.71	41104.03	4743.926	86421.49	29125.83	29125.83	137550.2	10589.31	45477.46	139241	187612.7
Volume Flow cum/hr	28.357	147.627	149.366	7.274	21.105	19.954	2.095	1.975	12.118	41.762	4.156	93.274	31.719	29.593	149.141	10.958	48.322	143.623	175.75
Enthalpy Gcal/hr	-34.862	-539.682	-538.033	-12.736	-22.138	-22.519	-6.33	-6.486	-44.506	-155.341	-13.372	-316.474	-108.108	-110.068	-505.701	-38.776	-164.808	-524.132	-636.125
Mass Flow kg/hr																			
WATER	1934.448	143320	143320	1924.776	9.672	9.672	907.256	907.256	11807.67	41103.69	2839.908	84517.41	29123.86	29123.86	135496.1	10085.76	43573.44	139241	154875.5
ETHANOL	19610.08	0	0	3922.017	15688.07	15688.07	0	0	0.031	0.343	0	0.056	1.961	1.961	0.431	492.884	0.004	0	0
CARBO-01	5.011	0	0	0	5.011	5.011	0	0	0	0	0	0	0	0	0	10.666	0	0	0
GLUCOSE	0	0	0	0	0	0	1614.918	1614.918	0.001	0	1497.223	1497.223	0	0	1614.919	0	1497.223	0	32298.38
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	0	0	0	0	0	0	438.773	438.773	0	0	406.795	406.795	0	0	438.773	0	406.795	0	438.773
CELLULOS																			
PROTEIN-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

TABLE 26: ASPEN MODEL 3 SOLID STREAM TABLE 1

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
Mole Flow kmol/hr	0	362.987	362.987	0	0	0	200.006	200.006	0	0	0	0	0	0	200.006	0	0	0	200.006
Mass Flow kg/hr	0	31824	31824	0	0	0	5397.93	5397.93	0	0	0	0	0	0	5397.93	0	0	0	5397.93
Volume Flow cum/hr	0	9.806	9.806	0	0	0	5.403	5.403	0	0	0	0	0	0	5.403	0	0	0	5.403
Enthalpy Gcal/hr		-42.064	-42.019				-3.977	-4.129							-3.984				-4.08
Mass Flow kg/hr																			
WATER	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHANOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CARBO-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	0	26693	26693	0	0	0	266.93	266.93	0	0	0	0	0	0	266.93	0	0	0	266.93
CELLU-01	0	389	389	0	0	0	389	389	0	0	0	0	0	0	389	0	0	0	389
XYLAN	0	310	310	0	0	0	310	310	0	0	0	0	0	0	310	0	0	0	310
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	0	4432	4432	0	0	0	4432	4432	0	0	0	0	0	0	4432	0	0	0	4432
CELLULOS																			
PROTEIN-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

TABLE 27: ASPEN MODEL 3 LIQUID STREAM TABLE 2

	21	22	23	24	25	26	29	30	31	32	35	B1	B2	B3	BRAN	C1-ETH	D1	D2	D3
Temperature C	30	30	82.6	100	93.6	100	30	30	30	30	30	101.5	100	42.3	25	86	95.2	70	28.5
Pressure bar	1.5	1.5	0.5	1.013	0.8	1.013	1.013	1.013	1.013	1.5	1.013	1.076	1.013	1.013	1.013	1.5	1.013	1.013	1.013
Vapor Frac	0	0	1	0	1	0	0	0	0	0	0	0	0	0	0	0	1	0	1
Mole Flow kmol/hr	8795.388	8795.388	2261.055	2261.055	2272.737	2272.737	839.919	933.243	911.897	911.897	8795.388	7549.362	1616.663	570.786	53.112	9507.214	1957.852	533.159	354.202
Mass Flow kg/hr	187612.7	187612.7	40733.54	40733.54	40944.02	40944.02	17866.61	19851.79	16428.09	16428.09	187612.7	137550.2	29125.83	10589.31	956.82	182378.9	44828.65	21549.54	15233.78
Volume Flow cum/hr	170.813	170.813	133766.7	44.359	86631.23	44.588	16.418	18.242	16.609	16.609	170.81	149.139	31.718	10.958	0.963	199.151	59183.99	28.355	8768.633
Enthalpy Gcal/hr	-640.937	-640.937	-129.543	-151.2	-130.01	-151.981	-60.133	-66.815	-62.167	-62.167	-640.94	-505.703	-108.108	-38.777	-3.626	-638.664	-111.204	-34.862	-32.784
Mass Flow kg/hr																			
WATER	154875.5	154875.5	40733.53	40733.53	40943.97	40943.97	14491.67	16101.85	16428.09	16428.09	154875.5	135496.1	29123.86	10085.76	956.82	164629.7	29133.61	1934.448	245.789
ETHANOL	0	0	0.004	0.004	0.052	0.052	0	0	0	0	0	0.431	1.961	492.884	0	15690.46	15690.03	19610.08	1.991
CARBO-01	0	0	0	0	0	0	0	0	0	0	0	0	0	10.666	0	5.011	5.011	5.011	14986
GLUCOSE	32298.38	32298.38	0	0	0	0	2936.174	3262.415	0	0	32298.38	1614.919	0	0	0	1614.919	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	438.773	438.773	0	0	0	0	438.773	487.526	0	0	438.773	438.773	0	0	0	438.773	0	0	0
CELLULOS																			
PROTEIN-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

TABLE 28: ASPEN MODEL 3 SOLID STREAM TABLE 2

	21	22	23	24	25	26	29	30	31	32	35	B1	B2	B3	BRAN	C1-ETH	D1	D2	D3
Mole Flow kmol/hr	200.006	200.006	0	0	0	0	0	39.223	78.677	78.677	200.006	200.006	0	0	78.677	200.006	0	0	0
Mass Flow kg/hr	5397.93	5397.93	0	0	0	0	0	6329.686	9753.394	9753.394	5397.93	5397.93	0	0	9753.394	5397.93	0	0	0
Volume Flow cum/hr	5.403	5.403	0	0	0	0	0	1.06	2.125	2.125	5.403	5.403	0	0	2.125	5.403	0	0	0
Enthalpy Gcal/hr	-4.14	-4.14						-3.181	-7.778	-7.778	-4.14	-3.984			-7.782	-4.017			
Mass Flow kg/hr																			
WATER	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHANOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CARBO-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	266.93	266.93	0	0	0	0	0	29.658	2965.841	2965.841	266.93	266.93	0	0	2965.841	266.93	0	0	0
CELLU-01	389	389	0	0	0	0	0	3501.911	3501.911	3501.911	389	389	0	0	3501.911	389	0	0	0
XYLAN	310	310	0	0	0	0	0	2793.191	2793.191	2793.191	310	310	0	0	2793.191	310	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	4432	4432	0	0	0	0	0	4.925	492.45	492.45	4432	4432	0	0	492.45	4432	0	0	0
CELLULOS																			
PROTEIN-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

TABLE 29: ASPEN MODEL 3 LIQUID STREAM TABLE 3

	DDGS-C	DDGS-H	E1-E12	E2-F1	E3-WAST	E5-E10	E9-WATER	E10-SSF	E11-E9	E12-SEP2	ETH-C	ETH-W	ETH-W-CC	ETHANOL	EVAP-E4	GRAIN	LIQUE	LIQUIDS
Temperature C	35	104.7	35	101.5	35	37.9	45	30	100	86	35	30	30	70.1	100.1	25	60	100.1
Pressure bar	1.013	1.013	1.013	1.076	1	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013
Vapor Frac	0	0	0	0	0	0	0	0	0	0.003	0	0	0.038	0	1	0	0	0.003
Mole Flow kmol/hr	78.535	78.535	9631.499	7549.362	1616.663	8795.388	655.426	8795.388	655.426	9631.499	341.185	8951.943	9306.331	341.185	2281.609	226.419	7955.469	6999.161
Mass Flow kg/hr	2960.948	2960.948	185682.9	137550.2	29125.83	187612.7	11807.71	187612.7	11807.71	185682.9	15702.75	172377.5	187612.7	15702.75	41104.03	4079	169746.1	127525.5
Volume Flow cum/hr	1.975	2.095	208.086	149.139	29.592	172.205	12.117	170.81	12.858	1030.643	19.952	177.035	8992.541	21.103	69873.7	4.104	159.343	784.787
Enthalpy Gcal/hr	-6.486	-6.33	-656.508	-505.703	-110.069	-639.551	-44.506	-640.94	-43.829	-646.878	-22.519	-611.123	-643.693	-22.139	-130.397	-15.456	-576.001	-468.843
Mass Flow kg/hr																		
WATER	907.256	907.256	166176.8	135496.1	29123.86	154875.5	11807.67	154875.5	11807.67	166176.8	9.672	154617.5	154875.5	9.672	41103.69	4079	140383.9	125621.1
ETHANOL	0	0	17178.66	0.431	1.961	0	0.031	0	0.031	17178.66	15688.07	15445.33	15692.45	15688.07	0.343	0	0	0.399
CARBO-01	0	0	273.761	0	0	0	0	0	0	273.761	5.011	260.989	14991.01	5.011	0	0	0	0
GLUCOSE	1614.918	1614.918	1614.919	1614.919	0	32298.38	0.001	32298.38	0.001	1614.919	0	1614.919	1614.919	0	0	0	29362.21	1497.223
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	438.773	438.773	438.773	438.773	0	438.773	0	438.773	0	438.773	0	438.773	438.773	0	0	0	0	406.795
CELLULOS																		
PROTEIN-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

TABLE 30: ASPEN MODEL 3 SOLID STREAM TABLE 3

	DDGS-C	DDGS-H	E1-E12	E2-F1	E3-WAST	E5-E10	E9-WATER	E10-SSF	E11-E9	E12-SEP2	ETH-C	ETH-W	ETH-W-CC	ETHANOL	EVAP-E4	GRAIN	LIQUE	LIQUIDS
Mole Flow kmol/hr	200.006	200.006	200.006	200.006	0	200.006	0	200.006	0	200.006	0	200.006	200.006	0	0	362.987	200.006	0
Mass Flow kg/hr	5397.93	5397.93	5397.93	5397.93	0	5397.93	0	5397.93	0	5397.93	0	5397.93	5397.93	0	0	31824	5397.93	0
Volume Flow cum/hr	5.403	5.403	5.403	5.403	0	5.403	0	5.403	0	5.403	0	5.403	5.403	0	0	9.806	5.403	0
Enthalpy Gcal/hr	-4.129	-3.977	-4.129	-3.984		-4.123		-4.14		-4.018		-4.14	-4.14			-42.158	-4.074	
Mass Flow kg/hr																		
WATER	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHANOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CARBO-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	266.93	266.93	266.93	266.93	0	266.93	0	266.93	0	266.93	0	266.93	266.93	0	0	26693	266.93	0
CELLU-01	389	389	389	389	0	389	0	389	0	389	0	389	389	0	0	389	389	0
XYLAN	310	310	310	310	0	310	0	310	0	310	0	310	310	0	0	310	310	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	4432	4432	4432	4432	0	4432	0	4432	0	4432	0	4432	4432	0	0	4432	4432	0
CELLULOS																		
PROTEIN-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

TABLE 31: ASPEN MODEL 3 LIQUID STREAM TABLE 4

	M1-E1	M2-DRYER	RCYC-WET	SEP1-C3	SEP2-P1	SEP2-SE3	SEP3-C3	SEP3-M1	SOLIDS	SYRUP	W-E6	WASTEBR	WASTWA	WATER	WATER-B	WATER2	WVAP
Temperature C	31.3	96.7	70.1	30	86	86	70	70	100.1	82.6	50	30	100	25	25	25	104.7
Pressure bar	1.013	1.013	1.013	1.013	0.841	0.841	1.013	1.013	1.013	0.5	1.013	1.013	1.013	1.013	1.013	1.013	1.013
Vapor Frac	0	0	0	1	0	1	1	0	0.003	0	0	0	0.561	0	0	0	1
Mole Flow kmol/hr	9631.499	733.962	191.975	354.388	9507.214	124.285	15.515	108.77	550.201	183.76	7729.05	93.324	2281.609	7729.05	858.786	555.084	655.426
Mass Flow kg/hr	185682.9	14768.65	5846.793	15235.22	182378.9	3304.015	587.875	2716.141	10024.73	4743.926	139241	1985.179	41104.03	139241	15471.26	10000	11807.71
Volume Flow cum/hr	191.303	15.091	7.241	8815.506	199.145	4412.945	436.868	3.211	61.692	4.155	143.621	1.824	39196.97	140.088	15.565	10.061	20320.2
Enthalpy Gcal/hr	-657.177	-50.22	-12.736	-32.57	-638.669	-7.268	-1.099	-7.274	-36.856	-13.372	-524.134	-6.681	-140.139	-527.602	-58.623	-37.891	-37.434
Mass Flow kg/hr																	
WATER	166176.8	12714.93	1924.776	258.073	164629.7	1547.063	73.474	1473.589	9875.022	2839.908	139241	1610.185	41103.69	139241	15471.26	10000	11807.67
ETHANOL	17178.66	0.031	3922.017	247.118	15690.46	1488.202	247.757	1240.445	0.031	0	0	0	0.343	0	0	0	0.031
CARBO-01	273.761	0	0	14730.02	5.011	268.75	266.644	2.107	0	0	0	0	0	0	0	0	0
GLUCOSE	1614.919	1614.919	0	0	1614.919	0	0	0	117.696	1497.223	0	326.242	0	0	0	0	0.001
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	438.773	438.773	0	0	438.773	0	0	0	31.978	406.795	0	48.753	0	0	0	0	0
CELLULOS																	
PROTEIN-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

TABLE 32: ASPEN MODEL 3 SOLID STREAM TABLE 4

	M1-E1	M2-DRYER	RCYC-WET	SEP1-C3	SEP2-P1	SEP2-SE3	SEP3-C3	SEP3-M1	SOLIDS	SYRUP	W-E6	WASTEBR	WASTWA	WATER	WATER-B	WATER2	WVAP
Mole Flow kmol/hr	200.006	200.006	0	0	200.006	0	0	0	200.006	0	0	39.223	0	0	0	0	0
Mass Flow kg/hr	5397.93	5397.93	0	0	5397.93	0	0	0	5397.93	0	0	6329.686	0	0	0	0	0
Volume Flow cum/hr	5.403	5.403	0	0	5.403	0	0	0	5.403	0	0	1.06	0	0	0	0	0
Enthalpy Gcal/hr	-4.137	-3.994			-4.018				-3.987			-3.181					
Mass Flow kg/hr																	
WATER	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHANOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CARBO-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	266.93	266.93	0	0	266.93	0	0	0	266.93	0	0	29.658	0	0	0	0	0
CELLU-01	389	389	0	0	389	0	0	0	389	0	0	3501.911	0	0	0	0	0
XYLAN	310	310	0	0	310	0	0	0	310	0	0	2793.191	0	0	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	4432	4432	0	0	4432	0	0	0	4432	0	0	4.925	0	0	0	0	0
CELLULOS																	
PROTEIN-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

Model 4

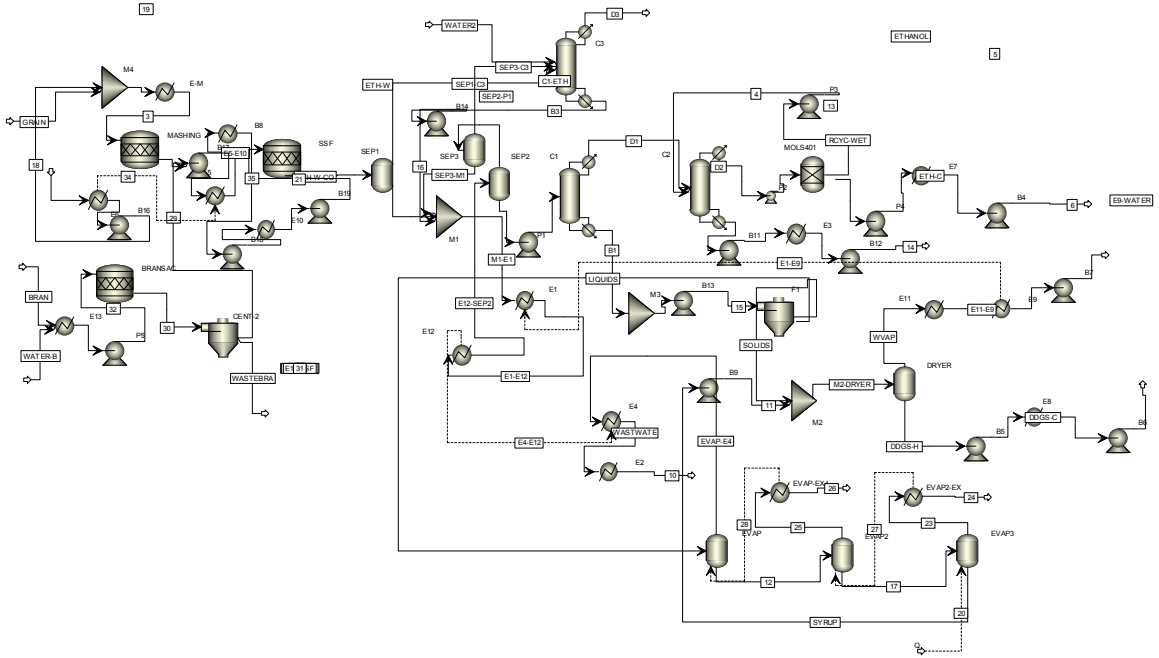

FIGURE 41: MODEL 4 ASPEN FLOW DIAGRAM

TABLE 33: ASPEN MODEL 4 LIQUID STREAM TABLE 1

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18
Temperature C	70	48.6	60	69.8	70.2	35	103.1	35	45	35	83	100.1	100	35	101.5	41.3	93.6	50
Pressure bar	1.5	1.013	1.013	1.5	1.5	1.5	1.5	1.5	1.5	1.013	1.5	1.013	1.5	1.5	1.5	1.5	0.8	1.5
Vapor Frac	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mole Flow kmol/hr	476.384	6562.096	6562.096	171.511	304.873	304.873	98.133	98.133	853.604	2007.689	131.604	4121.62	1444.528	1444.528	6949.442	569.62	2121.41	6375.643
Mass Flow kg/hr	19255.41	118218	118218	5223.873	14031.53	14031.53	3167.22	3167.22	15377.95	36169.3	3605.069	75486.35	26024.64	26024.64	126595.8	10535.08	39451.99	114859
Volume Flow cum/hr	25.338	121.76	123.205	6.499	18.859	17.83	2.496	2.351	15.782	36.748	3.071	81.499	28.342	26.442	137.293	10.877	41.942	118.474
Enthalpy Gcal/hr	-31.15	-445.168	-443.798	-11.378	-19.783	-20.123	-7.342	-7.512	-57.963	-136.692	-9.548	-276.309	-96.597	-98.349	-465.303	-38.711	-142.825	-432.353
Mass Flow kg/hr																		
WATER	1728.029	118218	118218	1719.389	8.64	8.64	1247.605	1247.605	15377.9	36168.94	1911.996	73793.22	26022.89	26022.89	124675.7	10086.1	37758.91	114859
ETHANOL	17522.42	0	0	3504.484	14017.94	14017.94	0	0	0.056	0.361	0	0.059	1.752	1.752	0.476	438.185	0.005	0
CARBO-01	4.958	0	0	0	4.958	4.958	0	0	0	0	0	0	0	0	0	10.796	0	0
GLUCOSE	0	0	0	0	0	0	1442.93	1442.93	0	0	1272.644	1272.644	0	0	1442.93	0	1272.644	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	0	0	0	0	0	0	476.685	476.685	0	0	420.429	420.429	0	0	476.685	0	420.429	0
CELLULOS																		
PROTEIN-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

TABLE 34: ASPEN MODEL 4 SOLID STREAM TABLE 1

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18
Mass Flow kg/hr	0	31099.47	31099.47	0	0	0	8044.671	8044.671	0	0	0	0	0	0	8044.671	0	0	0
Volume Flow cum/hr	0	9.165	9.165	0	0	0	5.324	5.324	0	0	0	0	0	0	5.324	0	0	0
Enthalpy Gcal/hr		-39.053	-39.011				-5.814	-5.961							-5.818			
Mass Flow kg/hr																		
WATER	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHANOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CARBO-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	0	24015.42	24015.42	0	0	0	960.617	960.617	0	0	0	0	0	0	960.617	0	0	0
CELLU-01	0	1760	1760	0	0	0	1760	1760	0	0	0	0	0	0	1760	0	0	0
XYLAN	0	1404	1404	0	0	0	1404	1404	0	0	0	0	0	0	1404	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	0	3920.055	3920.055	0	0	0	3920.055	3920.055	0	0	0	0	0	0	3920.055	0	0	0
CELLULOS																		
PROTEIN-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

TABLE 35: ASPEN MODEL 4 LIQUID STREAM TABLE 2

	19	21	22	23	24	25	26	29	30	31	32	35	B1	B2	B3	BRAN	C1-ETH	D1	D2
Temperature C	57.5	30	30	83	100	93.6	100	60	60	60	60	30	101.5	100	41.3	25	86	95.2	70
Pressure bar	1.5	1.5	1.5	0.5	1.013	0.8	1.013	1.013	1.013	1.013	1.5	1.013	1.076	1.013	1.013	1.013	1.5	1.013	1.013
Vapor Frac	0	0	0	1	0	1	0	0	0	0	0	0	0	0	0	0	0	1	0
Mole Flow kmol/hr	8003.212	8003.212	8003.212	1989.806	1989.806	2000.209	2000.209	1441.115	1601.239	1578.049	1578.049	8003.212	6949.442	1444.528	569.62	66.444	8698.847	1749.406	476.384
Mass Flow kg/hr	170253.6	170253.6	170253.6	35846.92	35846.92	36034.36	36034.36	28980.81	32200.9	28429	28429	170253.6	126595.8	26024.64	10535.08	1197	166652	40056.25	19255.41
Volume Flow cum/hr	159.826	155.297	155.297	117828.1	39.037	76242.78	39.241	28.359	31.51	29.628	29.629	155.294	137.291	28.341	10.877	1.204	181.913	52882.75	25.336
Enthalpy Gcal/hr	-577.93	-582.343	-582.343	-113.997	-133.061	-114.42	-133.757	-100.604	-111.782	-106.724	-106.724	-582.345	-465.305	-96.598	-38.711	-4.536	-584.169	-99.365	-31.15
Mass Flow kg/hr																			
WATER	140918.3	140918.3	140918.3	35846.92	35846.92	36034.31	36034.31	25261.89	28068.76	28429	28429	140918.3	124675.7	26022.89	10086.1	1197	150707.3	26031.61	1728.029
ETHANOL	0	0	0	0.005	0.005	0.054	0.054	0	0	0	0	0	0.476	1.752	438.185	0	14020.16	14019.69	17522.42
CARBO-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	10.796	0	4.958	4.958	4.958
GLUCOSE	28858.6	28858.6	28858.6	0	0	0	0	3242.24	3602.489	0	0	28858.6	1442.93	0	0	0	1442.93	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	476.685	476.685	476.685	0	0	0	0	476.685	529.65	0	0	476.685	476.685	0	0	0	476.685	0	0
CELLULOS																			
PROTEIN-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

TABLE 36: ASPEN MODEL 4 SOLID STREAM TABLE 2

	19	21	22	23	24	25	26	29	30	31	32	35	B1	B2	B3	BRAN	C1-ETH	D1	D2
Mass Flow kg/hr	8044.671	8044.671	8044.671	0	0	0	0	0	3202.1	6974	6974	8044.671	8044.671	0	0	6974	8044.671	0	0
Volume Flow cum/hr	5.324	5.324	5.324	0	0	0	0	0	0.539	1.706	1.706	5.324	5.324	0	0	1.706	5.324	0	0
Enthalpy Gcal/hr	-5.912	-5.971	-5.971						-1.62	-6.676	-6.676	-5.971	-5.818			-6.7	-5.851		
Mass Flow kg/hr																			
WATER	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHANOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CARBO-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	960.617	960.617	960.617	0	0	0	0	0	32.75	3275	3275	960.617	960.617	0	0	3275	960.617	0	0
CELLU-01	1760	1760	1760	0	0	0	0	0	1760	1760	1760	1760	1760	0	0	1760	1760	0	0
XYLAN	1404	1404	1404	0	0	0	0	0	1404	1404	1404	1404	1404	0	0	1404	1404	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	3920.055	3920.055	3920.055	0	0	0	0	0	5.35	535	535	3920.055	3920.055	0	0	535	3920.055	0	0
CELLULOS																			
PROTEIN-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

TABLE 37: ASPEN MODEL 4 LIQUID STREAM TABLE 3

	D3	DDGS-C	DDGS-H	E1-E12	E2-F1	E3-WAST	E5-E10	E9-WATER	E10-SSF	E11-E9	E12-SEP2	ETH-C	ETH-W	ETH-W-CC	ETHANOL	EVAP-E4	GRAIN	LIQUE	LIQUIDS
Temperature C	27.9	35	103.1	36.6	101.5	35	40	45	30	100	86	35	30	30	70.1	100.1	25	57	100
Pressure bar	1.013	1.013	1.013	1.013	1.076	1	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013	1.013
Vapor Frac	1	0	0	0	0	0	0	0	0	0	0.003	0	0	0.037	0	1	0	0	0.003
Mole Flow kmol/hr	315.977	98.133	98.133	8802.453	6949.442	1444.528	8003.212	853.604	8003.212	853.604	8802.453	304.873	8143.234	8459.739	304.873	2007.689	186.453	6562.096	6129.308
Mass Flow kg/hr	13601.6	3167.22	3167.22	169403.9	126595.8	26024.64	170253.6	15377.95	170253.6	15377.95	169403.9	14031.53	156647.3	170253.6	14031.53	36169.3	3359	141272.8	111655.7
Volume Flow cum/hr	7805.269	2.351	2.496	195.981	137.291	26.441	156.897	15.781	155.294	16.746	918.701	17.829	160.853	8034.018	18.857	61484.75	3.379	131.447	680.066
Enthalpy Gcal/hr	-29.265	-7.512	-7.342	-599.553	-465.305	-98.349	-580.753	-57.963	-582.345	-57.082	-591.019	-20.123	-555.716	-584.807	-19.783	-114.743	-12.728	-477.327	-410.388
Mass Flow kg/hr																			
WATER	211.046	1247.605	1247.605	151997.8	124675.7	26022.89	140918.3	15377.9	140918.3	15377.9	151997.8	8.64	140687.7	140918.3	8.64	36168.94	3359	115656.4	109962.2
ETHANOL	1.039	0	0	15237.64	0.476	1.752	0	0.056	0	0.056	15237.64	14017.94	13803.66	14021.2	14017.94	0.361	0	0	0.419
CARBO-01	13389.51	0	0	248.825	0	0	0	0	0	0	248.825	4.958	236.327	13394.47	4.958	0	0	0	0
GLUCOSE	0	1442.93	1442.93	1442.93	1442.93	0	28858.6	0	28858.6	0	1442.93	0	1442.93	1442.93	0	0	0	25616.36	1272.644
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	0	476.685	476.685	476.685	476.685	0	476.685	0	476.685	0	476.685	0	476.685	476.685	0	0	0	0	420.429
CELLULOS																			
PROTEIN-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

TABLE 38: ASPEN MODEL 4 SOLID STREAM TABLE 3

	D3	DDGS-C	DDGS-H	E1-E12	E2-F1	E3-WAST	E5-E10	E9-WATER	E10-SSF	E11-E9	E12-SEP2	ETH-C	ETH-W	ETH-W-CC	ETHANOL	EVAP-E4	GRAIN	LIQUE	LIQUIDS
Mass Flow kg/hr	0	8044.671	8044.671	8044.671	8044.671	0	8044.671	0	8044.671	0	8044.671	0	8044.671	8044.671	0	0	31099.47	8044.671	0
Volume Flow cum/hr	0	5.324	5.324	5.324	5.324	0	5.324	0	5.324	0	5.324	0	5.324	5.324	0	0	9.165	5.324	0
Enthalpy Gcal/hr		-5.961	-5.814	-5.957	-5.818		-5.95		-5.971		-5.851		-5.971	-5.971			-39.141	-5.913	
Mass Flow kg/hr																			
WATER	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHANOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CARBO-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	0	960.617	960.617	960.617	960.617	0	960.617	0	960.617	0	960.617	0	960.617	960.617	0	0	24015.42	960.617	0
CELLU-01	0	1760	1760	1760	1760	0	1760	0	1760	0	1760	0	1760	1760	0	0	1760	1760	0
XYLAN	0	1404	1404	1404	1404	0	1404	0	1404	0	1404	0	1404	1404	0	0	1404	1404	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	0	3920.055	3920.055	3920.055	3920.055	0	3920.055	0	3920.055	0	3920.055	0	3920.055	3920.055	0	0	3920.055	3920.055	0
CELLULOS																			
PROTEIN-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

TABLE 39: ASPEN MODEL 4 LIQUID STREAM TABLE 4

	M1-E1	M2-DRYER	RCYC-WET	SEP1-C3	SEP2-P1	SEP2-SE3	SEP3-C3	SEP3-M1	SOLIDS	SYRUP	W-E6	WASTEBR	WASTWA	WATER	WATER-B	WATER2	WVAP
Temperature C	31.2	98.7	70.1	30	86	86	70	70	100	83	50	60	100	25	25	25	103.1
Pressure bar	1.013	1.013	1.013	1.013	0.841	0.841	1.013	1.013	1.013	0.5	1.013	1.013	1.013	1.013	1.013	1.013	1.013
Vapor Frac	0	0	0	1	0	1	1	0	0.003	0	0	0	0.557	0	0	0	1
Mole Flow kmol/hr	8802.453	951.737	171.511	316.505	8698.847	103.606	14.007	89.599	820.133	131.604	6375.643	160.124	2007.689	6375.643	1511.606	555.084	853.604
Mass Flow kg/hr	169403.9	18545.17	5223.873	13606.3	166652	2751.918	530.373	2221.545	14940.1	3605.069	114859	3220.09	36169.3	114859	27232	10000	15377.95
Volume Flow cum/hr	174.515	19.347	6.469	7873.166	181.908	3678.696	394.409	2.621	90.996	3.071	118.472	3.151	34273.86	115.557	27.398	10.061	26354.95
Enthalpy Gcal/hr	-600.423	-64.457	-11.378	-29.091	-584.173	-6.076	-0.993	-5.994	-54.912	-9.549	-432.355	-11.178	-123.383	-435.215	-103.186	-37.891	-48.764
Mass Flow kg/hr																	
WATER	151997.8	16625.5	1719.389	230.618	150707.3	1290.573	66.526	1224.047	14713.51	1911.996	114859	2806.876	36168.94	114859	27232	10000	15377.9
ETHANOL	15237.64	0.056	3504.484	217.542	14020.16	1217.477	221.681	995.796	0.056	0	0	0	0.361	0	0	0	0.056
CARBO-01	248.825	0	0	13158.14	4.958	243.867	242.166	1.702	0	0	0	0	0	0	0	0	0
GLUCOSE	1442.93	1442.93	0	0	1442.93	0	0	0	170.286	1272.644	0	360.249	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLU-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	476.685	476.685	0	0	476.685	0	0	0	56.256	420.429	0	52.965	0	0	0	0	0
CELLULOS																	
PROTEIN-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

TABLE 40: ASPEN MODEL 4 SOLID STREAM TABLE 4

	M1-E1	M2-DRYER	RCYC-WET	SEP1-C3	SEP2-P1	SEP2-SE3	SEP3-C3	SEP3-M1	SOLIDS	SYRUP	W-E6	WASTEBR	WASTWA	WATER	WATER-B	WATER2	WVAP
Mass Flow kg/hr	8044.671	8044.671	0	0	8044.671	0	0	0	8044.671	0	0	3202.1	0	0	0	0	0
Volume Flow cum/hr	5.324	5.324	0	0	5.324	0	0	0	5.324	0	0	0.539	0	0	0	0	0
Enthalpy Gcal/hr	-5.969	-5.824			-5.851				-5.821			-1.62					
Mass Flow kg/hr																	
WATER	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHANOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CARBO-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
STARCH	960.617	960.617	0	0	960.617	0	0	0	960.617	0	0	32.75	0	0	0	0	0
CELLU-01	1760	1760	0	0	1760	0	0	0	1760	0	0	1760	0	0	0	0	0
XYLAN	1404	1404	0	0	1404	0	0	0	1404	0	0	1404	0	0	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
YEAST	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMMON-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CALCI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ENZYME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PROTEIN	3920.055	3920.055	0	0	3920.055	0	0	0	3920.055	0	0	5.35	0	0	0	0	0
CELLULOS																	
PROTEIN-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

9. APPENDIX B: SUBSIDY GRAPHS

In this section extra subsidy graphs is given not presented in chapter 4 (Results).

Sensitivity Analyses subsidy graphs for straight line depreciated asset value method subsidies using SAFEX values

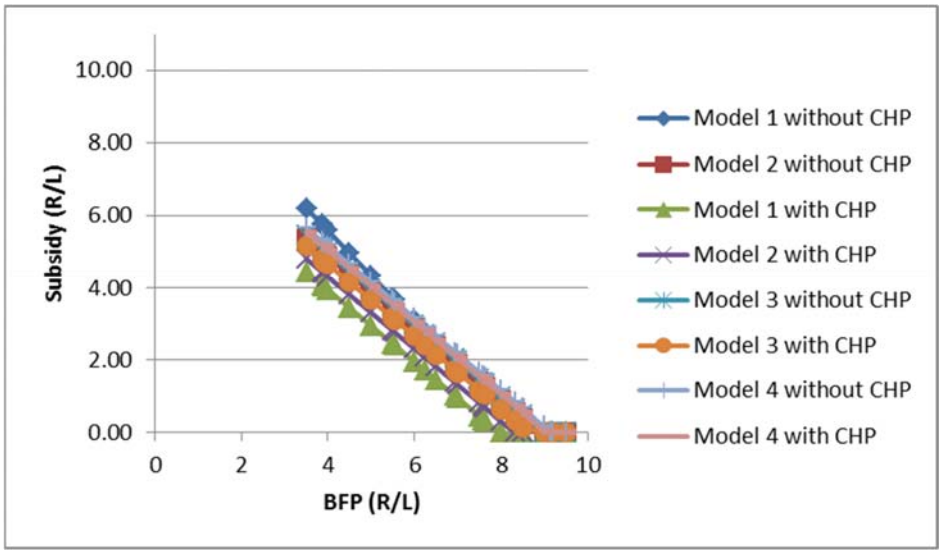

FIGURE 42: BFP VERSUS SUBSIDY FOR DEPRECIATED ASSET VALUE METHOD USING SAFEX VALUES

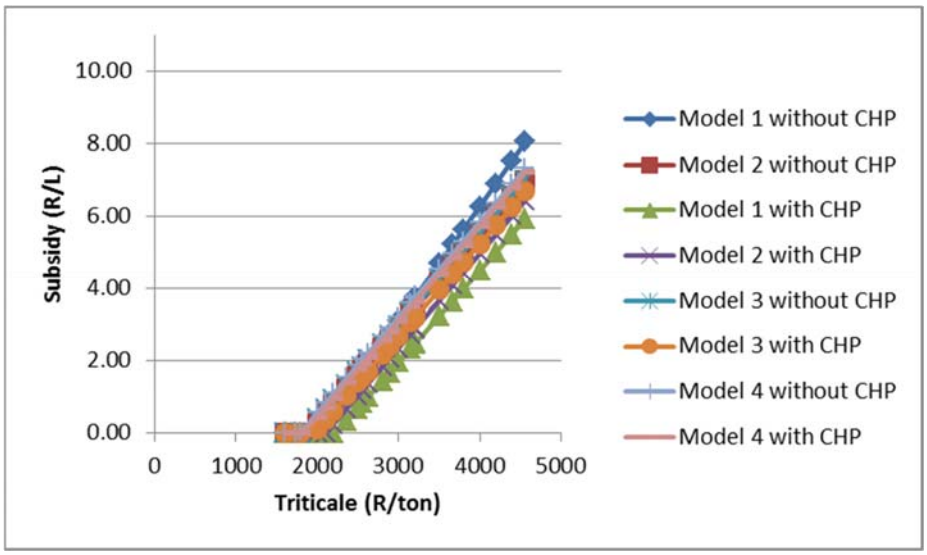

FIGURE 43: TRITICALE VERSUS SUBSIDY FOR DEPRECIATED ASSET VALUE METHOD USING SAFEX VALUES

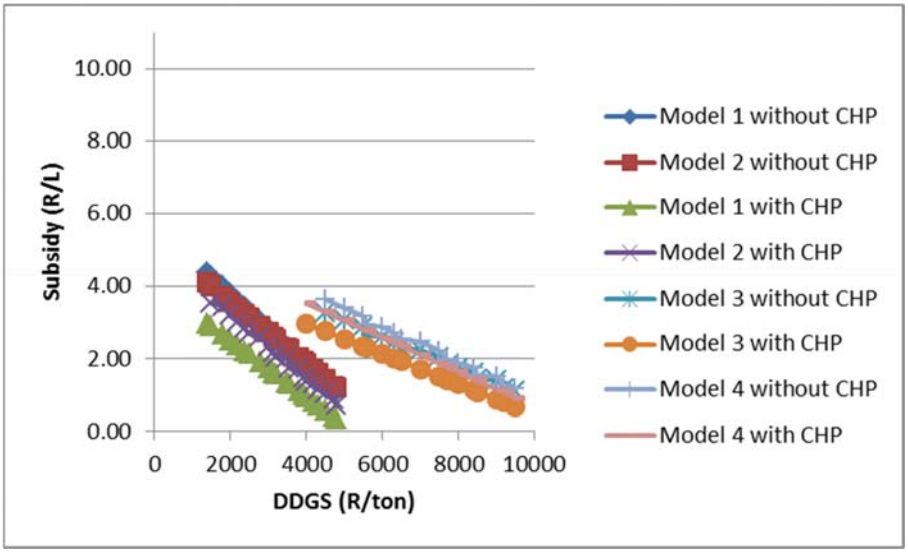

FIGURE 44: DDGS VERSUS SUBSIDY FOR DEPRECIATED ASSET VALUE METHOD USING SAFEX VALUES

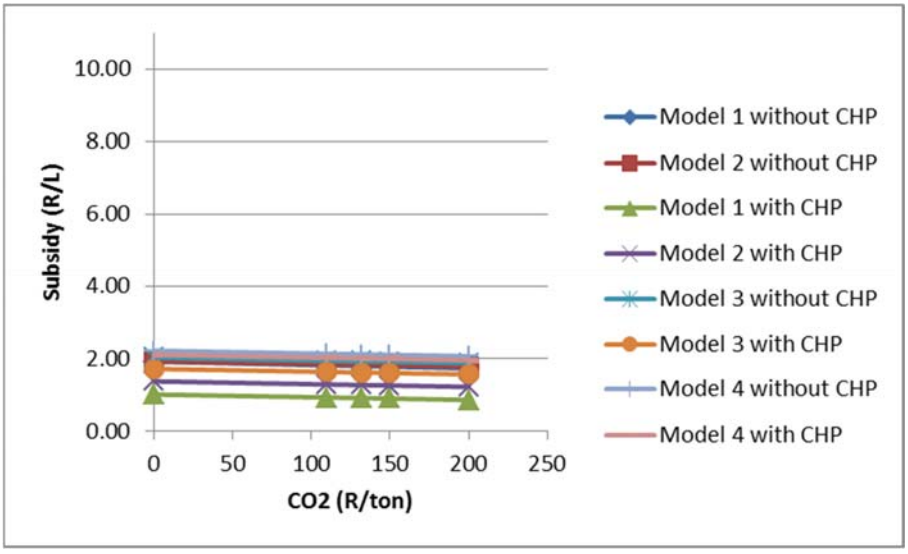

FIGURE 45: CO_2 VERSUS SUBSIDY FOR DEPRECIATED ASSET VALUE METHOD USING SAFEX VALUES

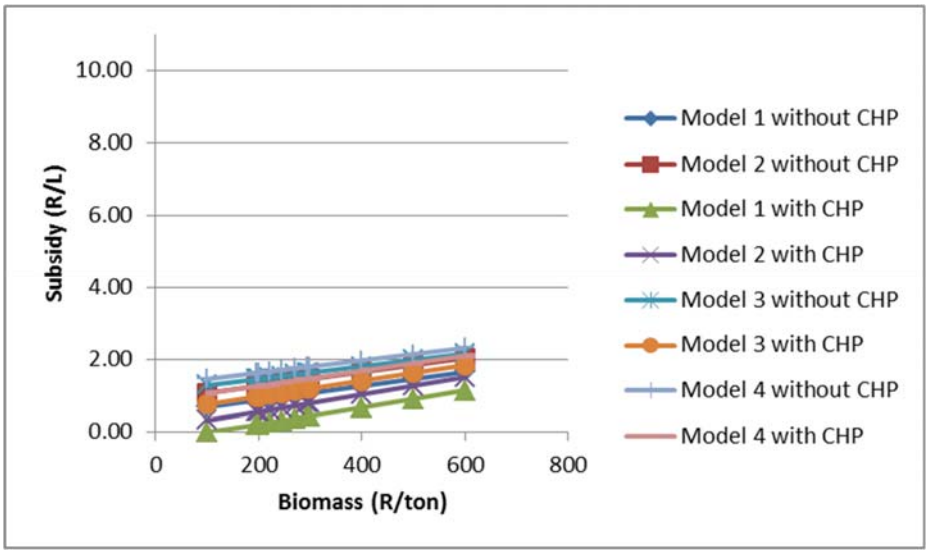

FIGURE 46: BIOMASS VERSUS SUBSIDY FOR DEPRECIATED ASSET VALUE METHOD USING SAFEX VALUES

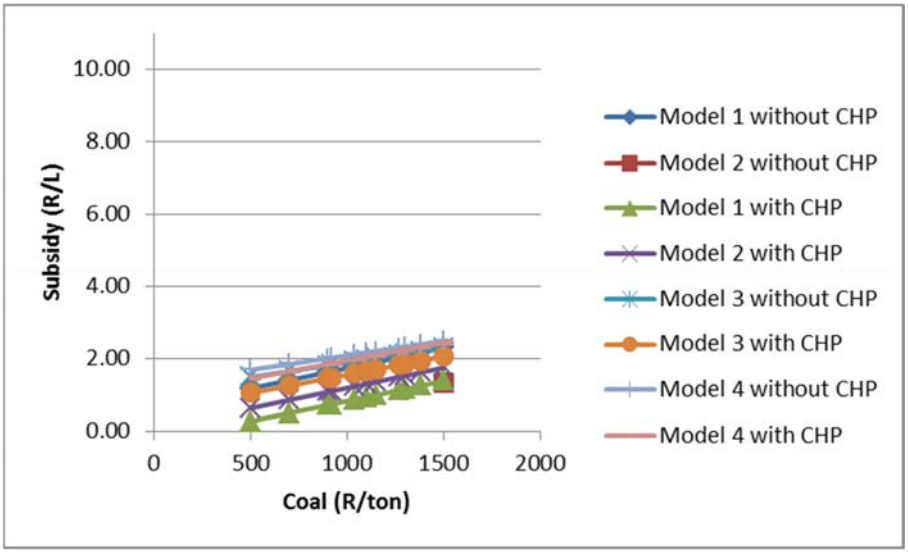

FIGURE 47: COAL VERSUS SUBSIDY FOR DEPRECIATED ASSET VALUE METHOD USING SAFEX VALUES

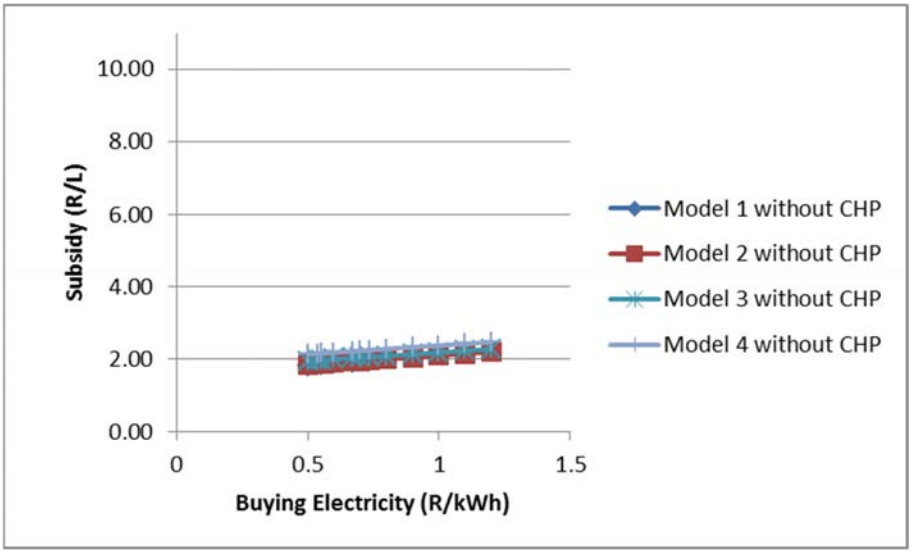

FIGURE 48: BUYING ELECTRICITY PRICE VERSUS SUBSIDY FOR DEPRECIATED ASSET VALUE METHOD USING SAFEX VALUES

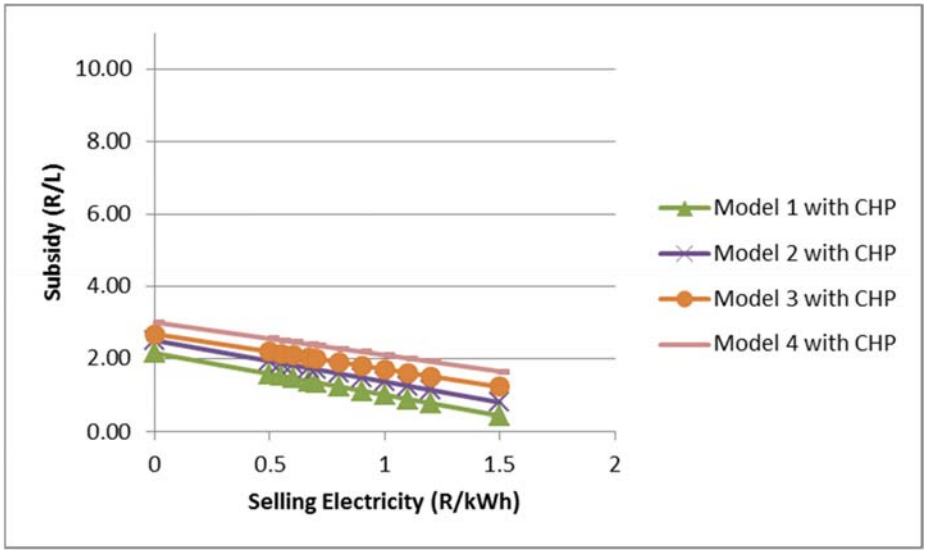

FIGURE 49: SELLING ELECTRICITY PRICE VERSUS SUBSIDY FOR DEPRECIATED ASSET VALUE METHOD USING SAFEX VALUES

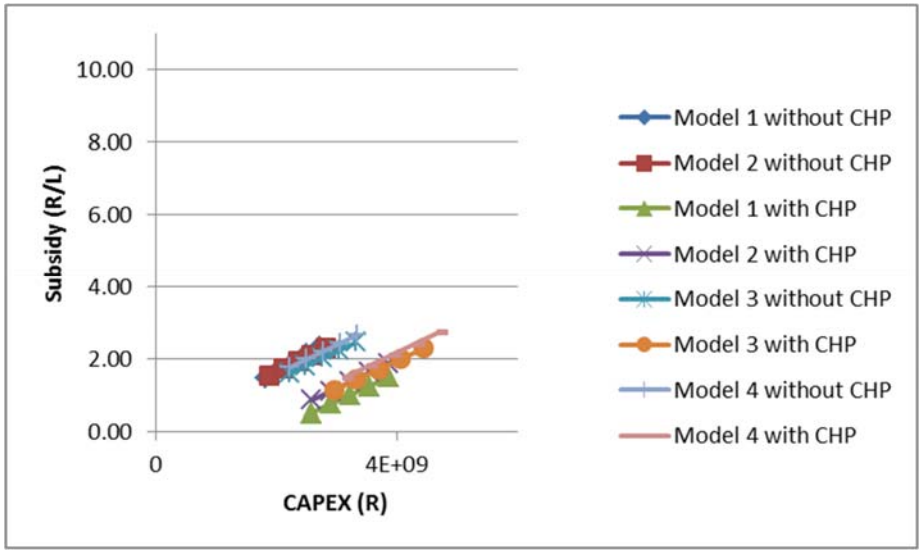

FIGURE 50: CAPEX VERSUS SUBSIDY FOR DEPRECIATED ASSET VALUE METHOD USING SAFEX VALUES

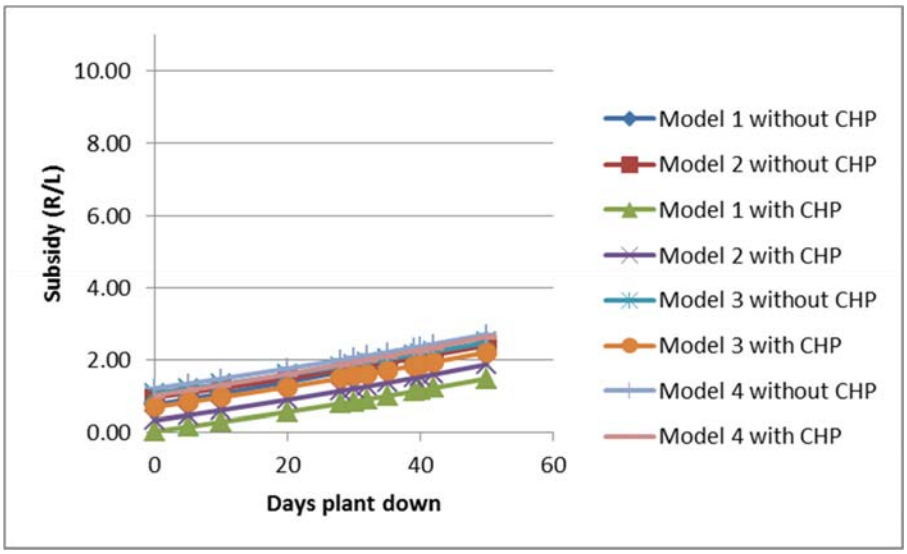

FIGURE 51: CAPACITY VERSUS SUBSIDY FOR DEPRECIATED ASSET VALUE METHOD USING SAFEX VALUES

Sensitivity Analyses subsidy graphs for straight line depreciated asset value method subsidy's using alternative values

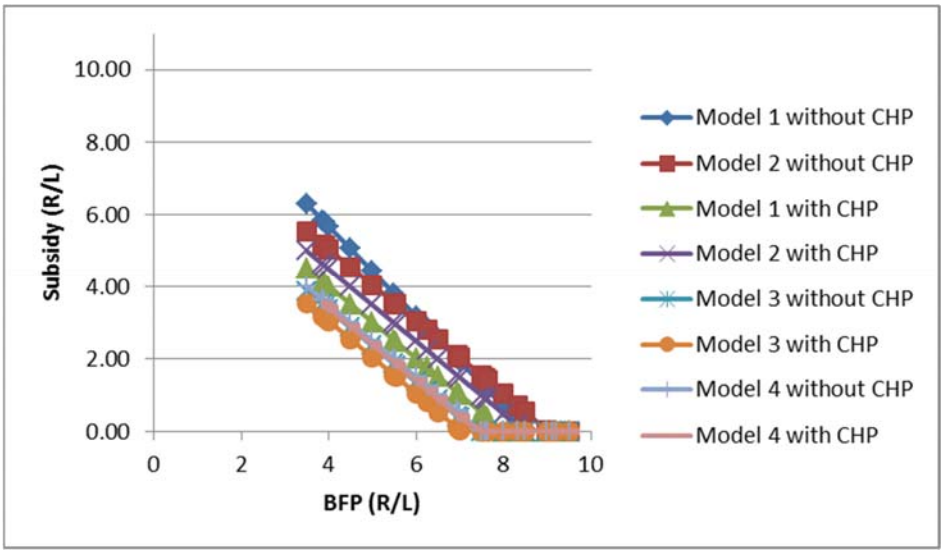

FIGURE 52: BFP VERSUS SUBSIDY FOR DEPRECIATED ASSET VALUE METHOD USING ALTERNATIVE VALUES

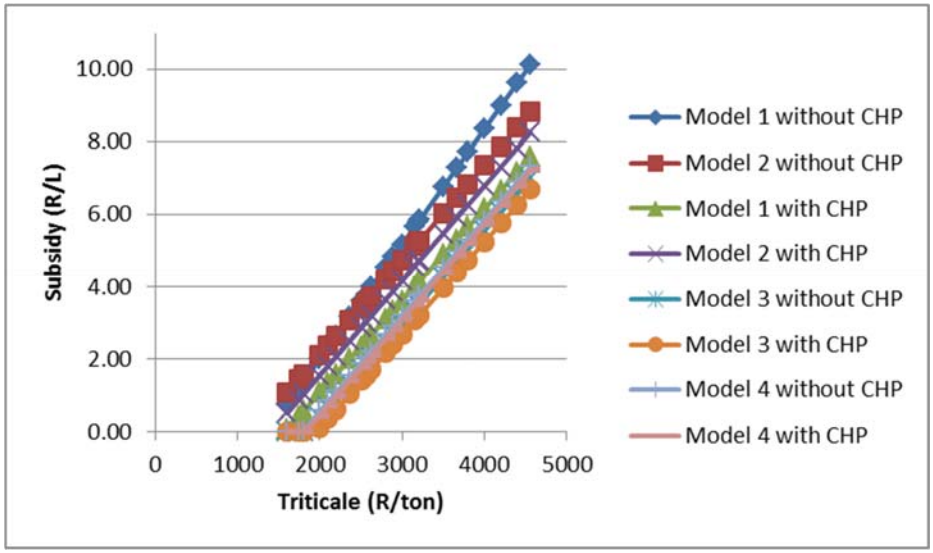

FIGURE 53: TRITICALE VERSUS SUBSIDY FOR DEPRECIATED ASSET VALUE METHOD USING ALTERNATIVE VALUES

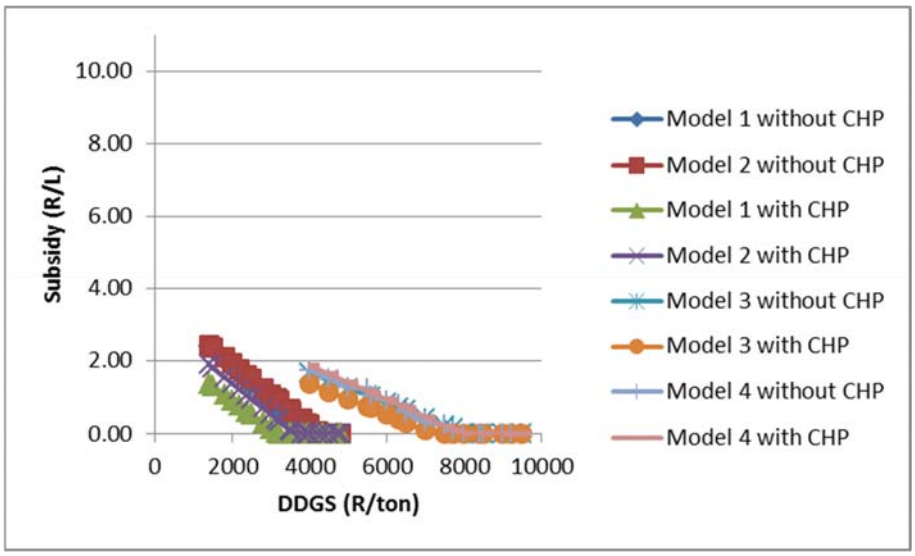

FIGURE 54: DDGS VERSUS SUBSIDY FOR DEPRECIATED ASSET VALUE METHOD USING ALTERNATIVE VALUES

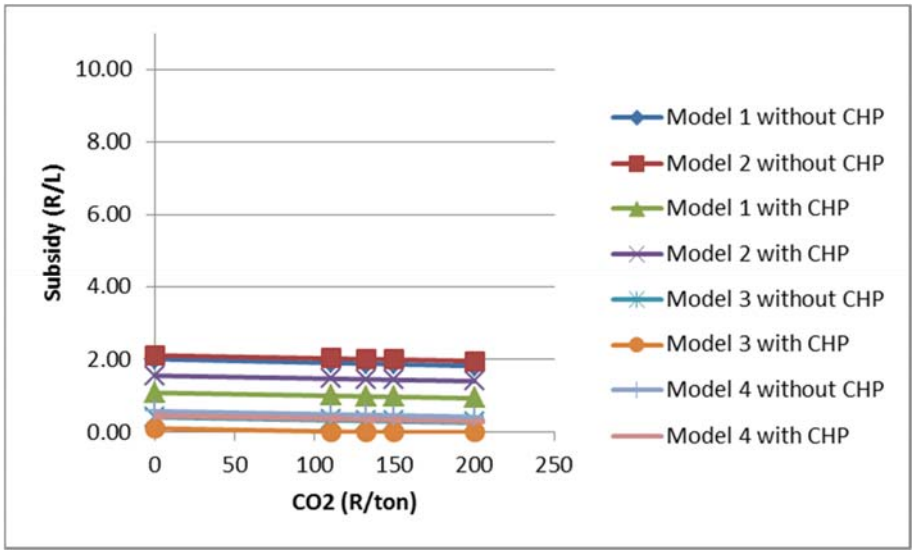

FIGURE 55: CO_2 VERSUS SUBSIDY FOR DEPRECIATED ASSET VALUE METHOD USING ALTERNATIVE VALUES

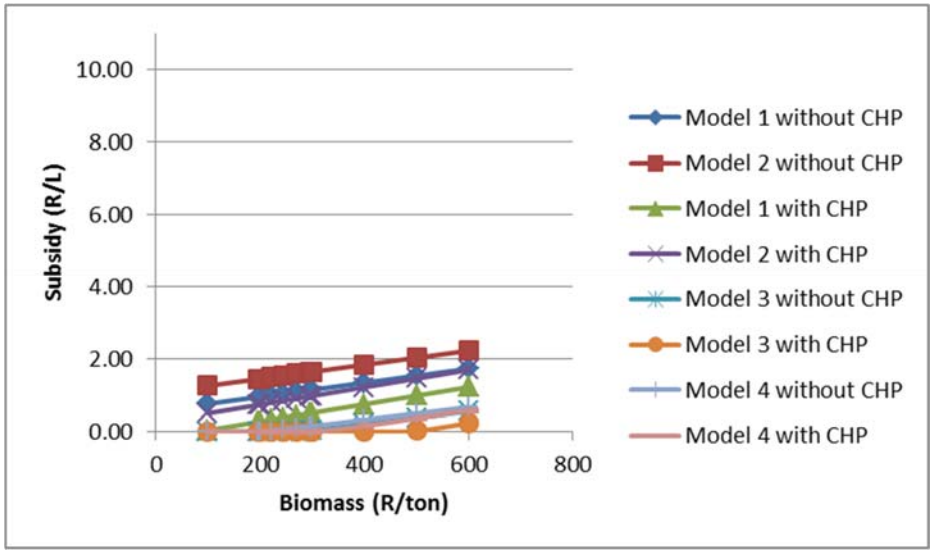

FIGURE 56: BIOMASS VERSUS SUBSIDY FOR DEPRECIATED ASSET VALUE METHOD USING ALTERNATIVE VALUES

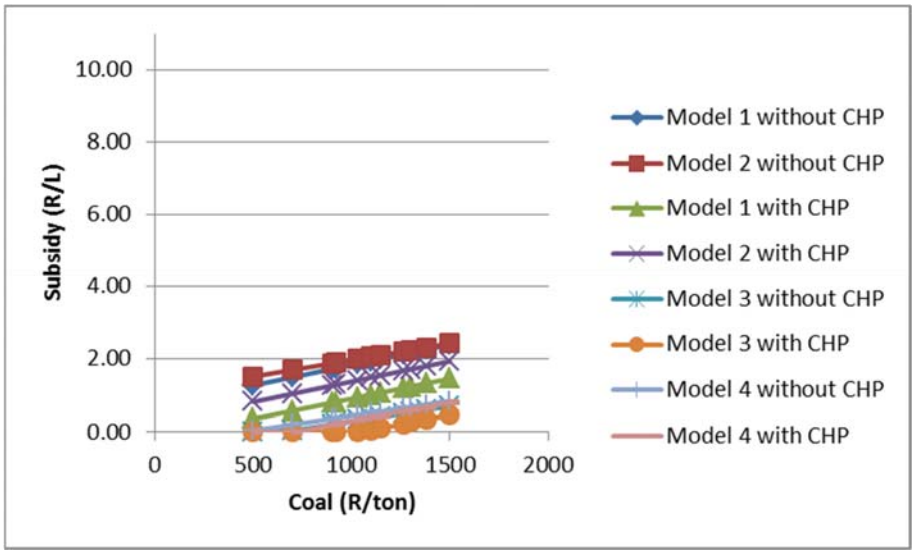

FIGURE 57: COAL VERSUS SUBSIDY FOR DEPRECIATED ASSET VALUE METHOD USING ALTERNATIVE VALUES

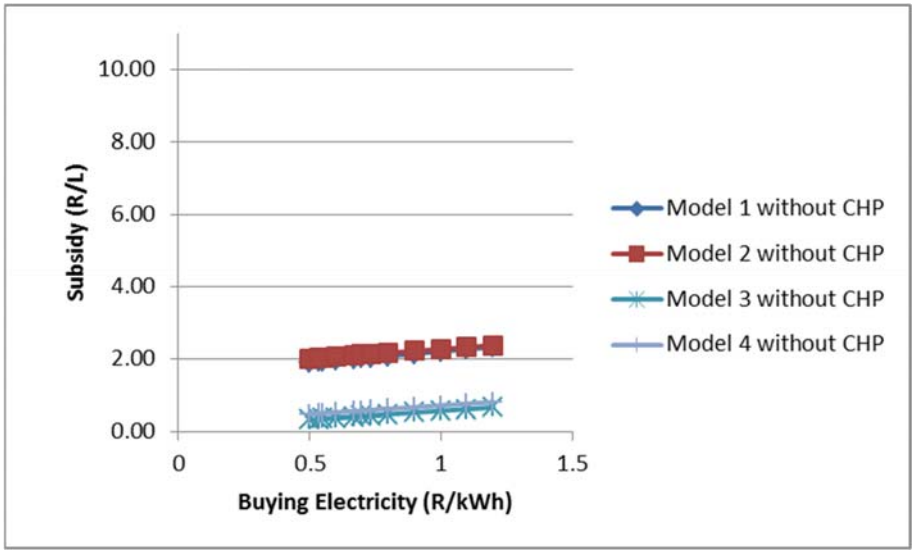

FIGURE 58: BUYING ELECTRICITY PRICE VERSUS SUBSIDY FOR DEPRECIATED ASSET VALUE METHOD USING ALTERNATIVE VALUES

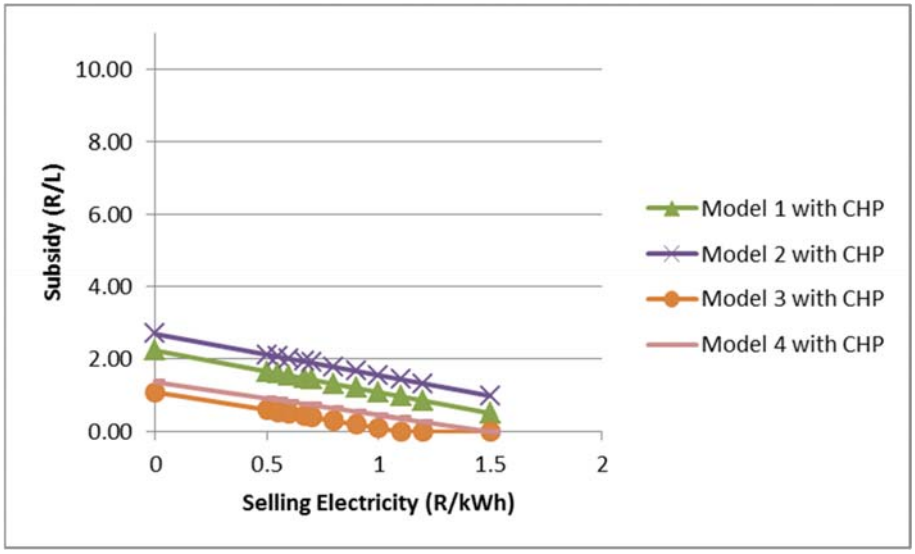

FIGURE 59: SELLING ELECTRICITY PRICE VERSUS SUBSIDY FOR DEPRECIATED ASSET VALUE METHOD USING ALTERNATIVE VALUES

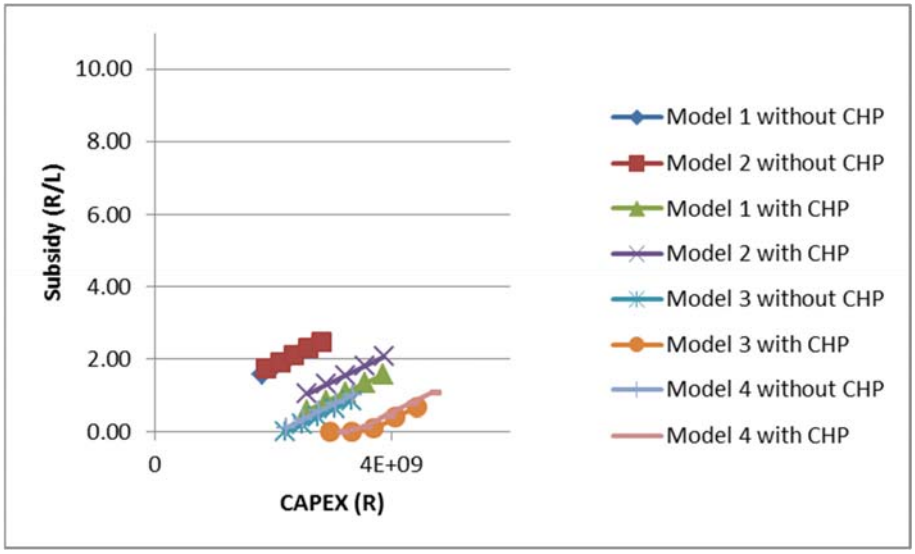

FIGURE 60: CAPEX VERSUS SUBSIDY FOR DEPRECIATED ASSET VALUE METHOD USING ALTERNATIVE VALUES

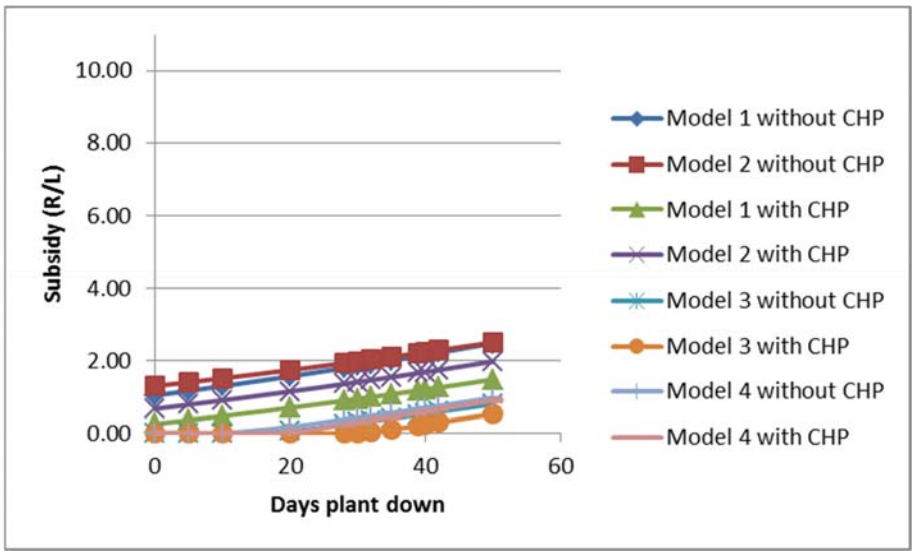

FIGURE 61: CAPACITY VERSUS SUBSIDY FOR DEPRECIATED ASSET VALUE METHOD USING ALTERNATIVE VALUES

Historical data Analyses subsidy graphs for depreciated asset value subsidy method.

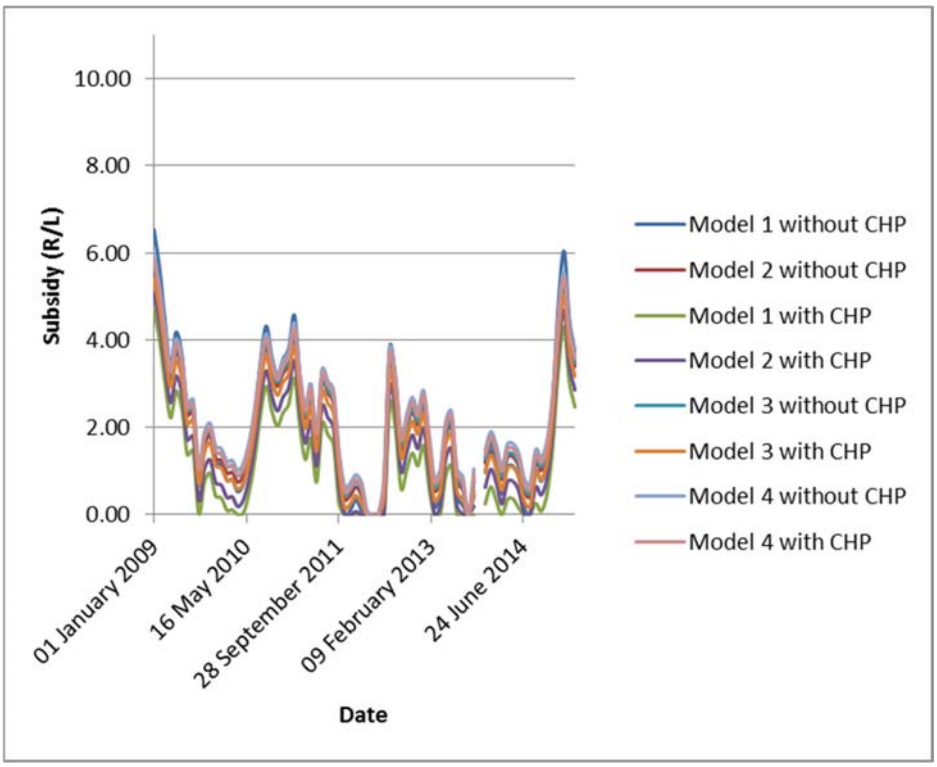

FIGURE 62: HISTORICAL DATA USED FOR SUBSIDY CALCULATION USING COAL FOR DEPRECIATED CAPEX SUBSIDY METHOD

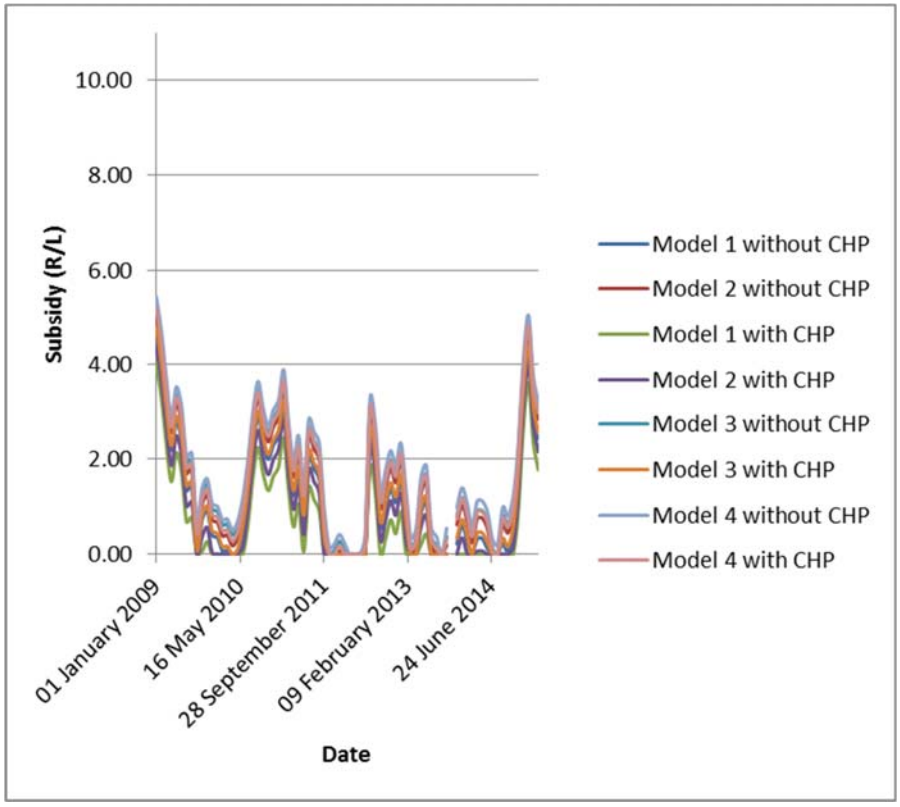

FIGURE 63: HISTORICAL DATA USED FOR SUBSIDY CALCULATION USING BIOMASS FOR DEPRECIATED ASSET VALUE SUBSIDY METHOD

10. APPENDIX C: IRR GRAPHS

In this section extra IRR graphs is given not presented in chapter 4 (Results).

Sensitivity Analyses IRR graphs for straight line depreciated asset value subsidy method using alternative values

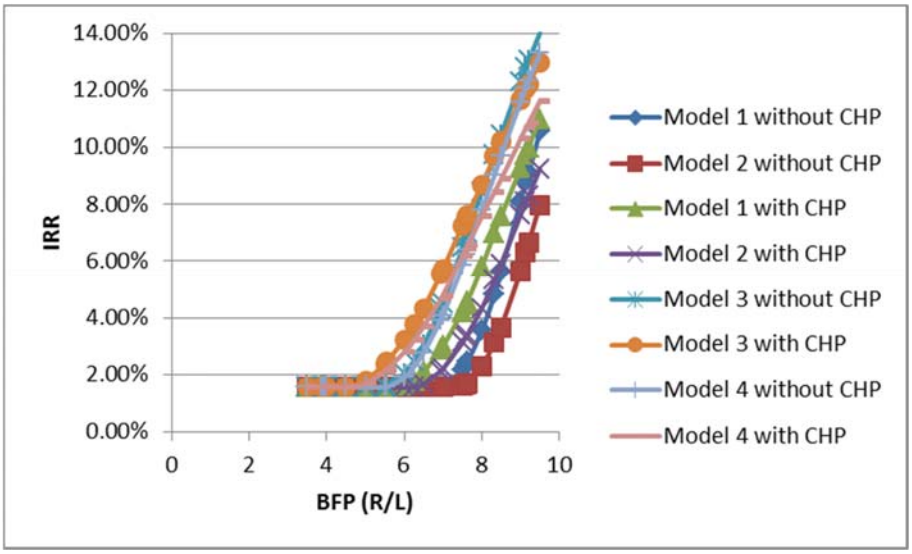

FIGURE 64: BFP VERSUS IRR FOR DEPRECIATED ASSET VALUE METHOD USING ALTERNATIVE VALUES

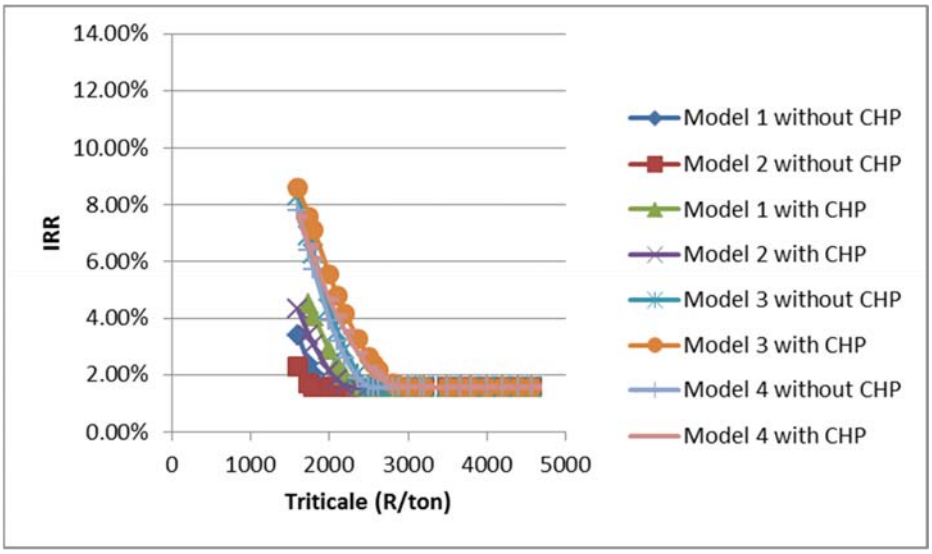

FIGURE 65: TRITICALE VERSUS IRR FOR DEPRECIATED ASSET VALUE METHOD USING ALTERNATIVE VALUES

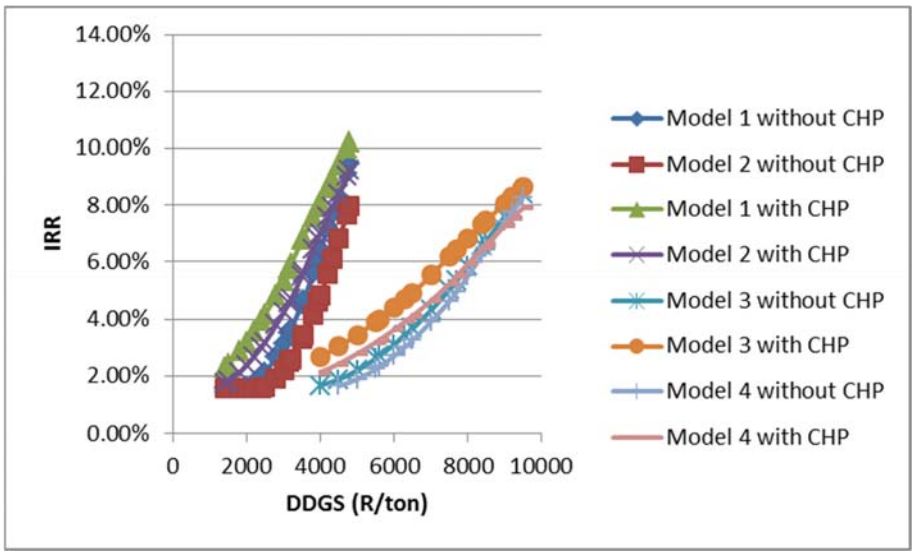

FIGURE 66: DDGS VERSUS IRR FOR DEPRECIATED ASSET VALUE METHOD USING ALTERNATIVE VALUES

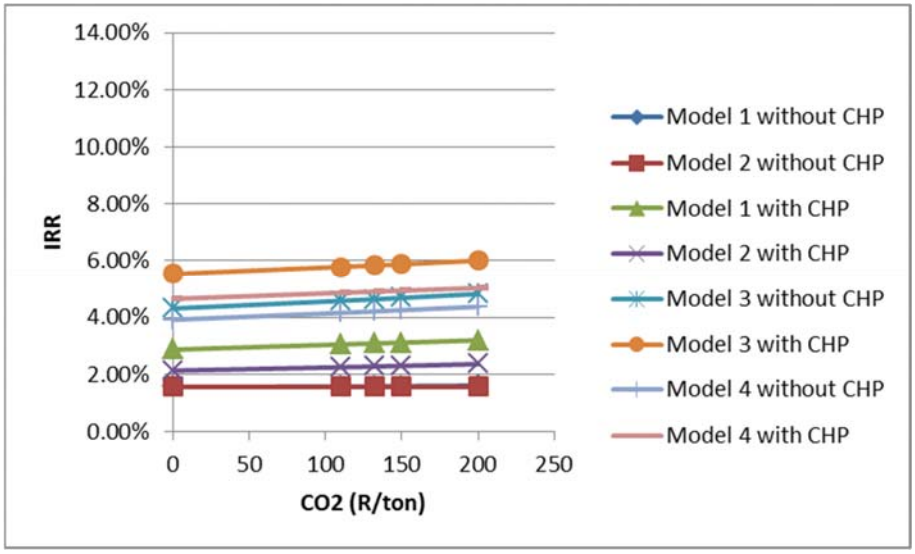

FIGURE 67: CO_2 VERSUS IRR FOR DEPRECIATED ASSET VALUE METHOD USING ALTERNATIVE VALUES

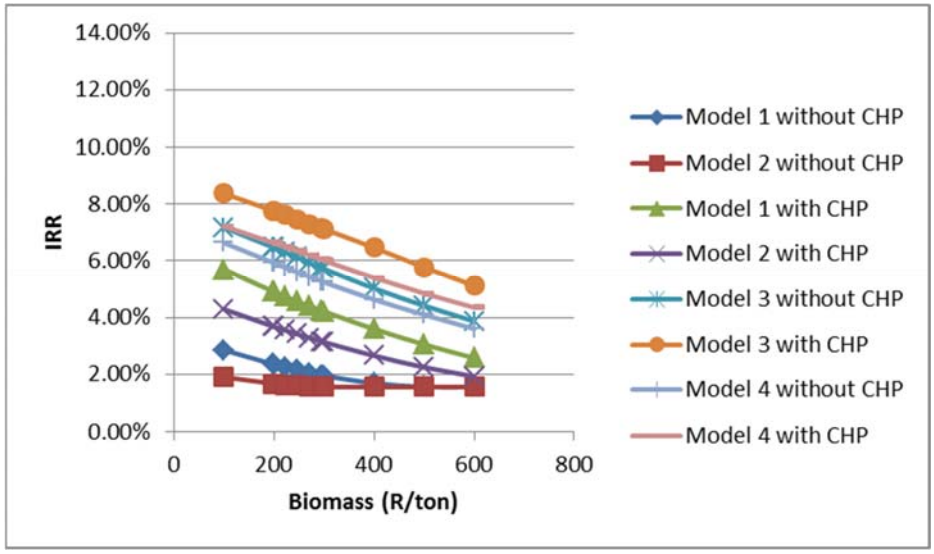

FIGURE 68: BIOMASS VERSUS IRR FOR DEPRECIATED ASSET VALUE METHOD USING ALTERNATIVE VALUES

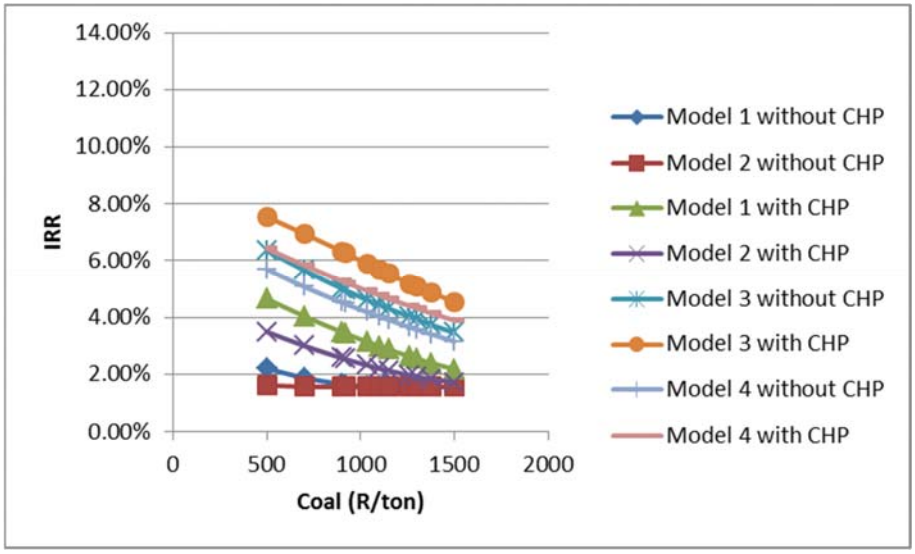

FIGURE 69: COAL VERSUS IRR FOR DEPRECIATED ASSET VALUE METHOD USING ALTERNATIVE VALUES

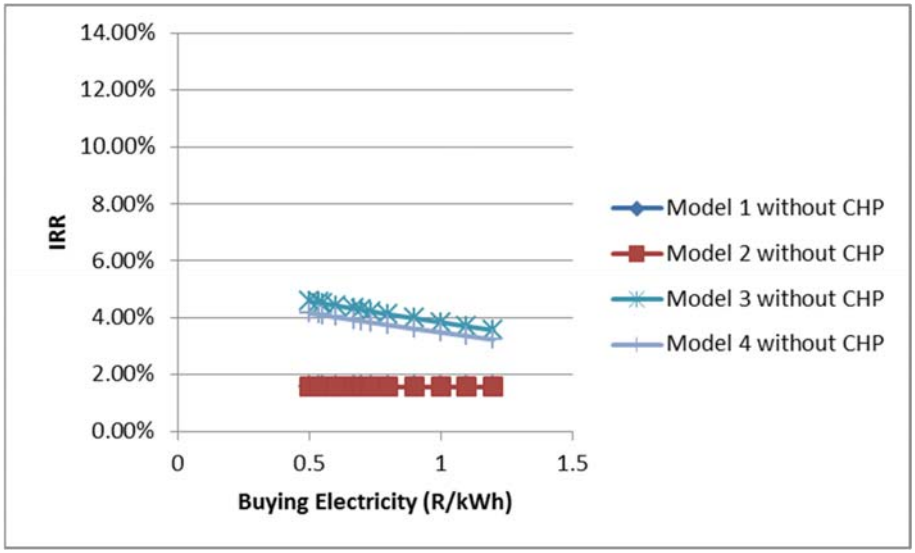

FIGURE 70: BUYING ELECTRICITY PRICE VERSUS IRR FOR DEPRECIATED ASSET VALUE METHOD USING ALTERNATIVE VALUES

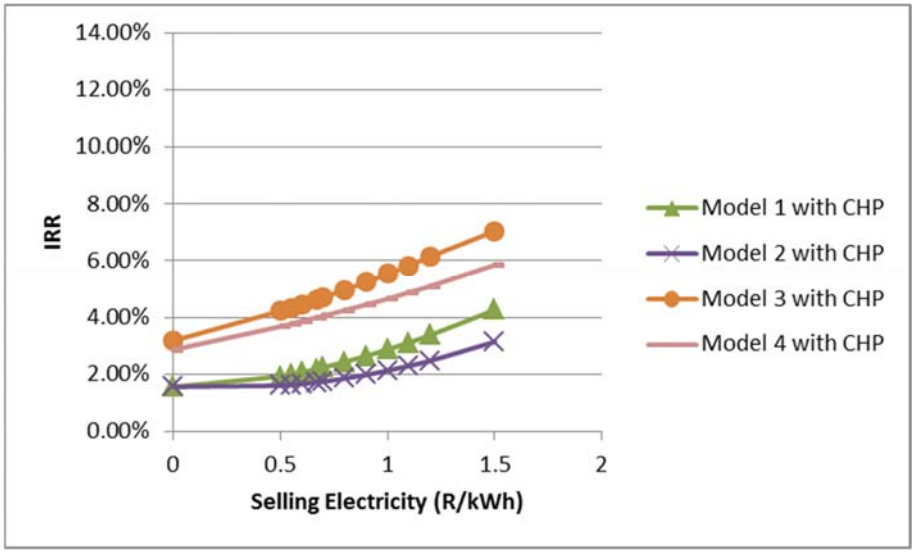

FIGURE 71: SELLING ELECTRICITY PRICE VERSUS IRR FOR DEPRECIATED ASSET VALUE METHOD USING ALTERNATIVE VALUES

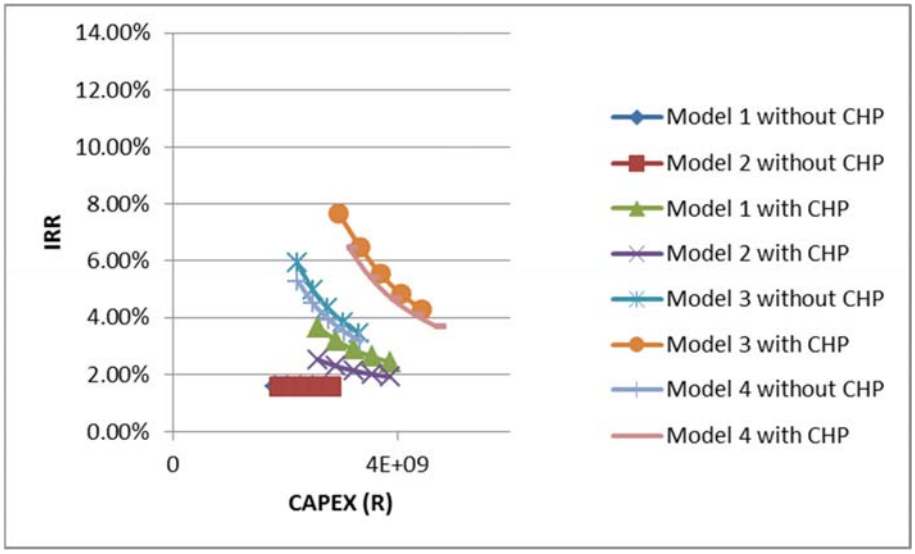

FIGURE 72: CAPEX ELECTRICITY PRICE VERSUS IRR FOR DEPRECIATED ASSET VALUE METHOD USING ALTERNATIVE VALUES

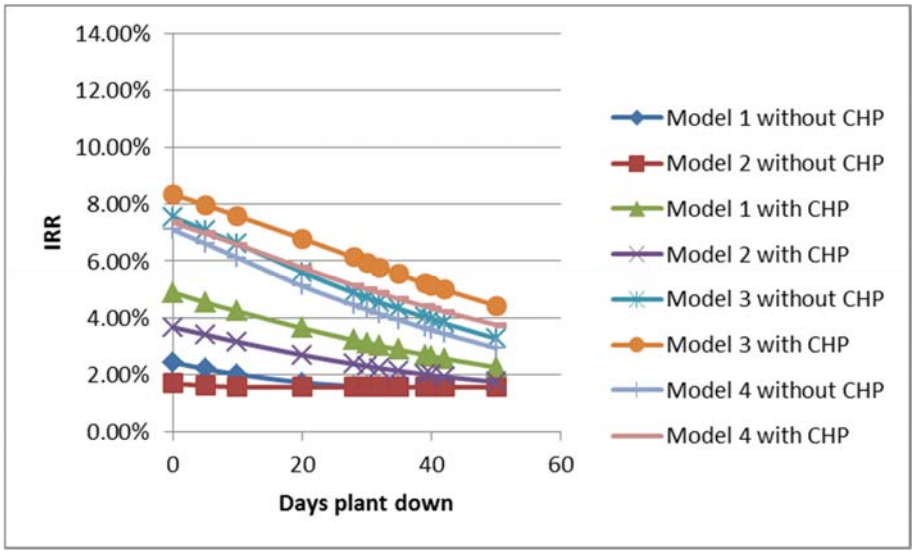

FIGURE 73: CAPACITY VERSUS IRR FOR DEPRECIATED ASSET VALUE METHOD USING ALTERNATIVE VALUES

Historical data Analyses graphs for depreciated asset value subsidy method.

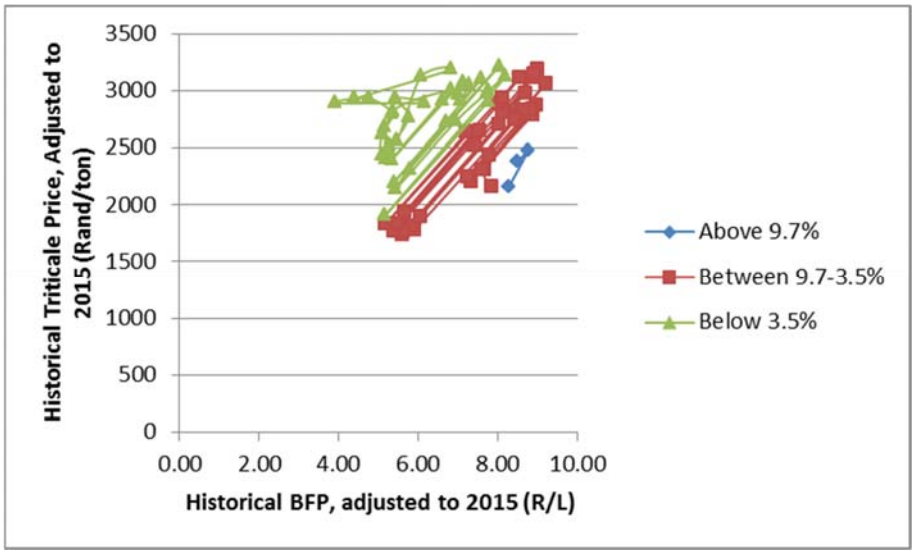

FIGURE 74: RELATIONSHIP BETWEEN TRITICALE AND BFP FOR DIFFERENT IRR GROUPS FOR COAL

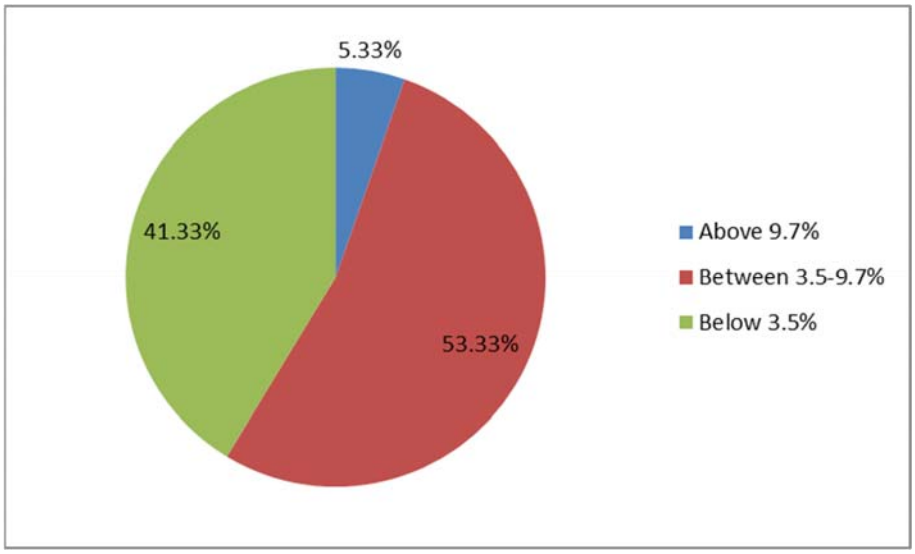

FIGURE 75: IRR RANGES FOR MODEL 1 WITH CHP USING BIOMASS